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ADVANCES IN TREATMENT OF DOMESTIC WASTES

CLEAN
WATER

TRAINING MANUAL

ENVIRONMENTAL PROTECTION AGENCY
OFFICE OF WATER PROGRAMS

REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
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SPRINGFIELD, VA. 22161

ADVANCES IN TREATMENT OF DOMESTIC WASTES

This course is offered for administrative, professional and technical personnel concerned with selection design and operation of wastewater treatment upgrading operations.

ENVIRONMENTAL PROTECTION AGENCY
OFFICE OF WATER PROGRAMS
TRAINING PROGRAM

1<

January 1972

FOREWORD

These manuals are prepared for reference use of students enrolled in scheduled training courses of the Office of Water Programs, Environmental Protection Agency.

> *Due to the limited availability of the manuals, it is not appropriate to cite them as technical references in bibliographies or other forms of publication.*
>
> *References to products and manufacturers are for illustration only; such references do not imply product endorsement by the Office of Water Programs, Environmental Protection Agency.*

The reference outlines in this manual have been selected and developed with a goal of providing the student with a fund of the best available current information pertinent to the subject matter of the course. Individual instructors may provide additional material to cover special aspects of their own presentations.

This manual will be useful to anyone who has need for information on the subjects covered. However, it should be understood that the manual will have its greatest value as an adjunct to classroom presentations. The inherent advantages of classroom presentation is in the give-and-take discussions and exchange of information between and among students and the instructional staff.

Constructive suggestions for improvement in the coverage, content, and format of the manual are solicited and will be given full consideration.

Joseph Bahnick

Joseph Bahnick
Acting Chief
Direct Technical Training Branch
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TRAINING PROGRAM

Through the Water Quality Office, Environmental Protection Agency conducts programs of research, technical assistance, enforcement, and technical training for water pollution control.

Training is available at five installations of the Agency. These are: the National Training Center located at the Robert A. Taft Sanitary Engineering Center in Cincinnati, Ohio; the Robert S. Kerr Water Research Center, Ada, Oklahoma; the Southeast Water Laboratory, Athens, Georgia; the Pacific Northwest Water Laboratory, Corvallis, Oregon; and the Hudson-Delaware Basins Office, Edison, New Jersey.

The objectives of the Training Program are to provide specialized training in the field of water pollution control which will lead to rapid application of new research findings through updating of skills of technical and professional personnel, and to train new employees recruited from other professional or technical areas in the special skills required. Increasing attention is being given to development of special courses providing an overview of the nature, causes, prevention, and control of water pollution.

Scientists, engineers, and recognized authorities from other Agency programs, from other government agencies, universities, and industry supplement the training staff by serving as guest lecturers. Most training is conducted in the form of short-term courses of one or two weeks' duration. Subject matter includes selected practical features of plant operation and design, and water quality evaluation in field and laboratory. Specialized aspects and recent developments of sanitary engineering, chemistry, aquatic biology, microbiology, and field and laboratory techniques not generally available elsewhere, are included.

The primary role and responsibility of the States in the training of wastewater treatment plant operators are recognized. Technical support of operator-training programs of the States is available through technical consultations in the planning and development of operator-training courses. Guest appearances of instructors from the Environmental Protection Agency, and the loan of instructional materials such as lesson plans and visual training aids, may be obtained through special arrangement. These training aids, including reference training manuals, may be reproduced freely by the states for their own training programs. Special categories of training for personnel engaged in treatment plant operations may be developed and made available to the States for their own further production and presentation.

An annual Bulletin of Courses is prepared and distributed by the Water Quality Office. The Bulletin includes descriptions of courses, schedules, application blanks, and other appropriate information. Organizations and interested individuals not on the mailing list should request a copy from one of the training centers mentioned above.

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On October 4, 1971 at the San Francisco meeting of the Water Pollution Control Federation four Process Design Manuals were released by the Environmental Protection Agency Technology Transfer.

These included:

Upgrading Existing Wastewater Treatment Plants
    Contract No. 14-12-933, Roy F. Weston, Inc.,

Suspended Solids Removal
    Contract No. 14-12-930, Burns & Roe, Inc.

Carbon Adsorption
    Contract No. 14-12-928, Swindell-Dressler Co.

Phosphorus Removal
    Contract No. 14-12-936, Black and Veatch Construction Engineers

Recommendations contained in these or later releases by the Environmental Protection Agency supersede materials contained in this manual that are not consistent with these Design Manuals.

SECTION I CURRENT STATUS AND TRENDS

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CHAPTER 2

INVESTIGATIVE APPROACH

2.1 Examination of Need for Upgrading

The need for upgrading an existing wastewater treatment plant may arise for one or more of the following reasons:

1. Lack of proper plant operation and control.
2. Inadequate plant design.
3. Changes in wastewater flow or characteristics.
4. Changes in treatment requirements.

2.1.1 Lack of Proper Plant Operation and Control

One of the primary considerations in evaluating an overloaded plant is in the area of improper plant operation and control. An incorrectly operated or maintained plant will never be able to perform according to design. Therefore, no physical upgrading should be considered before the engineer is assured that the plant is being operated to yield its maximum efficiency.

There seem to be two main reasons for the large number of smaller-capacity plants throughout the country which are poorly operated or maintained:

1. The smaller community or sanitary district will not or cannot provide funds for the employment of qualified operators.

2. The lack of appropriation of operating funds limits the extent of any scheduled maintenance program.

In addition, many wastewater treatment plants do not have a laboratory equipped to analyze the wastewater samples from the various units to assess their performance. Improper operation, coupled with inadequate laboratory control, increases the probability of inadequate treatment. For this reason, the smaller community or sanitary district should make sure its plants are staffed with an adequate number of competent operators and laboratory personnel. Further, sufficient funds should be made available to insure a proper maintenance program.

2.1.2 Inadequate Plant Design

In the past, the problems associated with inadequately designed wastewater treatment plants have been a major concern of individual state agencies. For this reason, most states have adopted conservative design guidelines and review procedures which must be followed unless

This copy taken from the "Process Design Manual for Upgrading Existing Wastewater Treatment Plants" prepared by Roy F. Weston, Inc. under Contract 14-12-933 with the EPA Technology Transfer Program.

the engineer has operating data which will substantiate a less conservative viewpoint. The implementation of these procedures by regulatory agencies has substantially reduced the problem of inadequate plant design as applied to municipal wastewater.

The major exception to this is in the area of joint municipal and industrial wastewater treatment. Design of wastewater treatment facilities based strictly on state standards may not be applicable when a significant amount of industrial wastewater is discharged to a municipal plant. When this situation occurs, sufficient wastewater sampling and treatability studies should be performed to establish parameters necessary for the design of the treatment plant.

In the past, one of the areas in the design of treatment plants not given much consideration was flexibility. The following design considerations can greatly increase the flexibility allowed to the treatment plant operator:

1. Splitter boxes before and after individual unit processes, for greater ease in operation and maintenance.
2. Piping associated with the aeration basin designed with enough flexibility to facilitate implementation of various activated sludge modifications.
3. Sufficient blower capacity to meet fluctuating organic loads to the aeration basin.
4. Sufficient recycle capacity for trickling filters to meet fluctuating loads.
5. Chlorination capacity with an incremental factor to be utilized for operational control, e.g., odor control.

2.1.3 Changes in Wastewater Flow or Characteristics

Two major problems facing engineers in the design of wastewater treatment plants are: 1) forecasting changes in population and wastewater flow; and 2) the operational problems caused by changes in the characteristics of municipal wastewaters due to the rapid industrialization of an area. Forecasting changes in population and wastewater flows in connection with upgrading of a treatment plant may be quite burdensome, but generally will not be subject to as much uncertainty as in similar forecasting for a relatively undeveloped area. In many cases, the maximum anticipated growth is defined by saturation of the tributary area. Potential extension of this area must also be considered, and is often limited by topographical constraints and political boundaries.

In-line measurement and analysis of existing wastewater flows, analysis of local area growth patterns, examination of local influences such as land use planning studies, zoning regulations, wastewater discharge ordinances, and full use of State, County, and local planning agencies can all be extremely useful in judging the future expected flows for a given upgrading situation.

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Rapid industrialization in the plant's service area can cause major operational problems in existing plants, which may require upgrading of various unit processes to handle increased hydraulic and organic loads. The alert community before issuing a building permit should make the industry aware that pretreatment may be required for wastewater containing toxic materials, or for those having an unusually high percentage of organic or inorganic material compared to typical domestic wastewaters. Equalization of industrial wastewater discharges may be helpful in minimizing diurnal flow variations to treatment plants and in distributing shock loads of high-strength wastewaters.

Population equivalent is a reasonable method of equating the organic content or flow contribution of industries to the ordinary per capita contribution present in domestic wastewaters. In many cases, even a relatively small industry may contribute a significantly higher loading than the existing population. Population equivalents for many industrial wastewaters should be based on COD analysis rather than on BOD_5, since extremely strong or toxic wastes may show an artificially low BOD_5 value.

2.1.4 Changes in Treatment Requirements

Increased pressure on the part of the Federal and State governments and a more ecologically minded public are requiring local communities and sanitary districts to enforce existing water quality standards. In addition, many regulatory agencies are stipulating increased organic and solids removal, minimum dissolved oxygen concentration in the plant effluent, a minimum consistent BOD and SS removal during low-flow periods, and removal of substantial portions of the nitrogenous oxygen demand. For example, the Potomac Enforcement Conference has recently required 85 percent removal of nitrogenous oxygen demand from wastewaters treated in the Washington, D.C. metropolitan area (1). Therefore, to meet existing and future requirements, upgrading of existing treatment facilities will often be required.

2.2 Study of Plant Performance History

It is the responsibility of the engineer engaged in upgrading an existing wastewater treatment plant to examine the plant's performance history thoroughly, as the first step in an engineering study. Treatment plant operating records serve two basic functions: 1) providing the operating information necessary for process monitoring and control; and 2) providing the historical record of plant performance. Data collected from a typical plant should include: 1) daily total flow; 2) maximum and minimum flow rates within a day; 3) volume of air added; 4) a series of concentration parameters such as BOD, suspended solids, COD, phosphates, nitrogen, and temperature; 5) consumption of chlorine, coagulants, and neutralizing chemicals; and 6) sludge production.

The engineer should examine all treatment plant records to become familiar with the type of sampling and flow measurement techniques employed by the plant personnel and verify their accuracy. One difficult area in assessing plant performance data is the variation in reliability of influent flow measurement data. Often, plant flow is obtained from some type of flow recording instrument, and reliable flow information is possible only when the treatment plant operator makes it a point to calibrate the flow measuring and recording instruments periodically. A representative portion of the operating data should be evaluated by the engineer. If operational data are not available, then it is the responsibility of the engineer to collect sufficient data to proceed with his upgrading evaluation.

2.3 Identification of Problem Areas

After evaluating the plant's operating records, it is necessary to determine the factors which influence the plant's current performance. There are four problem areas whose effect should be considered in any upgrading situation:

1. Hydraulic and organic overloading.
2. Inadequate organic removal.
3. Inadequate solids removal.
4. Inadequate sludge handling.

The performance of different unit processes within a treatment plant is affected to varying degrees by an increase in hydraulic and organic loading. The relationship between the increases in flow and in organics is also an important consideration. For example, a significant increase in flow without a corresponding increase in organics will generally not be as detrimental as when the increase in flow is also accompanied by a correspondingly large increase in organics.

Nevertheless, the efficiency of most unit operations is affected by hydraulic and organic overloading. The increased flow will increase overflow rates and will decrease available detention time in primary and secondary clarifiers. An increased overflow rate in the primary clarifier will in turn decrease suspended solids and BOD removal at this unit. As a result, solids and BOD loads to secondary treatment processes are increased. Also, hydraulic overloading reduces the compaction of solids that normally takes place in clarifiers and increases the volume of sludge to be handled.

Organic overloading significantly increases the:

1. Organic load per unit of aeration volume.
2. Lbs. BOD applied per pound of MLVSS under aeration.
3. Demand for more oxygen.

These effects significantly decrease the efficiency of activated sludge treatment. A similar effect can be seen in trickling filter plants. In activated sludge plants, an increase in organic loading can reduce the operational stability of the process by causing sludge bulking. Solids removal efficiency in secondary clarifiers is thereby reduced, significantly minimizing the amount of solids that can be carried in the aeration basin. The carryover of biological solids from the final clarifier increases the effluent BOD.

Many treatment plants have anaerobic digestion facilities designed on the basis of a certain volume per capita. Increases in the number of people to be served, together with the increased volume of sludge resulting from poor clarifier performance caused by hydraulic overloading, makes the existing volume in many anaerobic digesters inadequate. This situation creates an operational problem leading to inadequate sludge digestion. The poorly digested sludge further complicates dewatering on vacuum filters or sand beds, because it does not dewater rapidly.

The above discussion clearly indicates that problem areas are interrelated and should be assessed concurrently to determine the logical combination of applicable upgrading procedures. Since optimization of existing facilities is necessary in any upgrading situation, an understanding of various modifications available for the different unit processes is essential. This is the reason for defining the capabilities and limitations of various unit processes in subsequent chapters of this manual.

2.4 Consideration of Applicable Upgrading Techniques

Technology in the field of wastewater treatment in the past decade has provided many innovative upgrading procedures to meet deficiencies in existing processes. Various research projects sponsored by EPA have resulted in a better understanding of various unit processes. In addition, new types of equipment for wastewater treatment have enlarged the range of alternatives available for consideration in upgrading treatment plants.

It has long been recognized that the performance of a wastewater treatment plant is affected by variations in the influent flow. Equalization of extreme flows can dampen the fluctuations in loading to a plant.

Various processes and process equipment are being marketed and successfully used to increase removal in primary and secondary clarifiers. They include the use of chemical coagulation, peripheral-feed clarifiers, and inclined-tube settlers. These procedures, in many cases, have the effect of maintaining good solids removal while maximizing the hydraulic throughput in the existing facilities. Chemical addition in primary and secondary clarifiers can increase solids capture and BOD removal. In addition, several types of screening devices are available as possible substitutes for primary clarification.

Several modifications of the conventional activated sludge process, including step aeration, contact stabilization, and complete mixing, have been adequately studied and have been used to upgrade various treatment plants. A most significant development in the activated sludge treatment process came with the full-scale demonstration of the feasibility and effectiveness of using oxygen aeration as a substitute for air aeration. Plants using oxygen aeration are now being designed, with the single largest plant, at Detroit, Michigan, designed for a flow of 300 mgd.

Another method of upgrading an overloaded secondary plant is to provide additional treatment ahead of the existing biological treatment facilities. The use of plastic media trickling filters should be considered when roughing treatment is indicated. Plastic media filters have been successfully used as roughing filters in industrial wastewater treatment, and it is very likely that they will be used in the future for upgrading of municipal treatment plants.

The true effect of a nitrified effluent on dissolved oxygen in receiving waters has just recently been recognized and substantiated (2). For this reason, some regulatory agencies are requiring nitrification of treatment plant effluents during summer periods, and in some cases are contemplating a year-round nitrification requirement. Nitrification during summer months may be accomplished through modifications to the existing treatment units, such as addition of chemicals to the primary clarifier to decrease the organic loading to existing aeration units. However, dependable year-round nitrification will require a two-stage biological treatment system.

On many occasions, treatment plants which are functioning satisfactorily (design flow not exceeded) are required to improve solids or BOD removal because of more stringent water quality standards. This additional treatment can often be achieved by polishing the treatment plant effluent. Several methods are currently available and have been used successfully, including aerobic and facultative lagoons, microstraining, multi-media filtration, and activated carbon treatment.

Although considerable effort has been made in the study of organic removal processes, the area of sludge handling and dewatering has not received corresponding attention. Inadequate digestion and sludge handling facilities often adversely affect the overall treatment plant operation. The return of supernatant or filtrate from thickening or dewatering units to the head of the plant can impose high oxygen demands on the system and add substantial amounts of fine solids which are difficult to remove in the secondary clarifier. The high concentration of nutrients and organics in such streams and the periodic nature of the return flow often necessitate separate treatment, especially when nutrient removal is a consideration.

Various sludge-handling developments which have been successful are: 1) high-rate anaerobic digestion; 2) aerobic digestion; 3) thickening of sludge prior to digestion to increase the capacity of existing digesters; 4) the use of chemicals to improve thickening and dewatering of sludges; and 5) the use of heat treatment processes for the disposal of sludges. The use of aerobic digestion is likely to alleviate many of the operational problems associated with the anaerobic treatment of sludges.

Having briefly presented a general overview of the technology available for upgrading, the engineer should keep in mind various considerations which will affect the overall economics of upgrading:

1. The physical condition of existing plant equipment and structures as it relates to the use of existing facilities in an upgrading situation.

2. The length of time before a major expansion will be required, based on population and wastewater flow projections.

3. The time required for implementation of various upgrading techniques.

4. Compatability of upgrading procedures with future planned expansions. For example, if the engineer determines that contact stabilization will not work well on a particular wastewater, then step aeration may not be the most economical interim step. Perhaps completely mixed activated sludge or oxygen aeration would be more logical. The reasoning behind this type of decision will be explained further in subsequent chapters.

5. Financial resources available to the community.

6. Costs of the various upgrading techniques that can be used to achieve essentially the same result. The operation and maintenance costs, as well as the capital costs, may be substantially different. Therefore, economic comparison of available alternatives is necessary.

2.5 References

1. *Nitrogen Removal from Wastewaters*. Federal Water Quality Administration, Publication ORD-17010, October, 1970.

2. Courchaine, Robert, *Significance of Nitrification in Stream Analysis - Effects on the Oxygen Balance*. Journal Water Pollution Control Federation, 40, No. 5, pp. 835-847 (1968).

STATUS OF ADVANCED WASTE TREATMENT

I The Federal Water Quality Administration is conducting an accelerated national program on advanced waste treatment research. Program objectives are to improve existing waste treatment processes and to develop and demonstrate practical means of removing maximum possible amounts of physical, chemical, and biological pollutants. Emphasis in advanced waste treatment studies is being placed both on treatment as a water pollution control measure and treatment for water reuse.

This discussion will cover the current status of Federal Water Quality Administration research on advanced waste treatment with emphasis on costs and design factors and processes for nutrient removal. Nutrient removal and processes for increased removal of BOD and suspended solids have high research and development program interest at this time.

A Research and development programs are conducted under four mechanisms:

1 In-house studies

2 Research grants to universities and nonprofit organizations

3 Research and development grants to cities, sanitary districts, and states, and

4 Research contracts

B Research contracts are awarded principally for determination of technical and economic feasibility of new and novel treatment techniques, and for extension of in-house studies. Most contract work scopes include laboratory studies, pilot-scale studies, and desktop analyses. In some cases, it is necessary to conduct larger scale studies to develop reliable design, operating cost, and maintenance cost data. If feasibility is established, a process is progressively evaluated in large-scale facilities to establish reliable design, performance, and cost data. The later phases of process development are usually conducted in Federal Water Quality Administration pilot plants or by award of cost-sharing Research and Development Grant Grants for pilot plant or full-scale plant studies.

During 1969 there were 93 advanced waste treatment research contracts in force with a total cost of 8.9 million. An additional 30 contracts were being processed at a cost of 2.2 million.

Forty-eight research and development grants were in force with a total cost of 25.7 million.

II SPECIFIC PROJECTS

A Processes for removal of suspended materials include sedimentation, screening coagulation-sedimentation, and filtration.

1 Microscreening equipment for removal of solids from secondary effluent is available employing a rotary drum configuration. A disadantage of this method is susceptibility to wide variations in flow and suspended solids. Seventy-five percent of suspended solids and sixty percent of BOD can be removed from secondary effluent by this method.

2 Filtration with sand or multi-media filters can be utilized. A higher degree of removal can be obtained by coagulation-sedimentation with or without filtration.

B Processes for removal of organic materials include biological, physical and chemical methods. Biodegradable organics may be removed by conventional biological treatment, with or without coagulation or filtration to remove particulates and by activated carbon adsorption. Removal of nonbiodegradable organics can be accomplished by activated carbon adsorption and chemical oxidation.

1 Extensive studies have been conducted on activated carbon treatment. Treatment by use of granular and powdered carbon has been evaluated. Granular carbon treatment is accomplished by contacting water with a fixed bed of carbon. Flow rates are normally 4 to 6 gallon per minute per square foot. Bed depth is designed to provide contact times of 20 to 40 minutes.

2 The use of powdered activated carbon for wastewater treatment is accomplished by slurry contacting. After contacting, the powdered carbon is settled out along with suspended solids. Filtration of settled effluent will produce a high quality water.

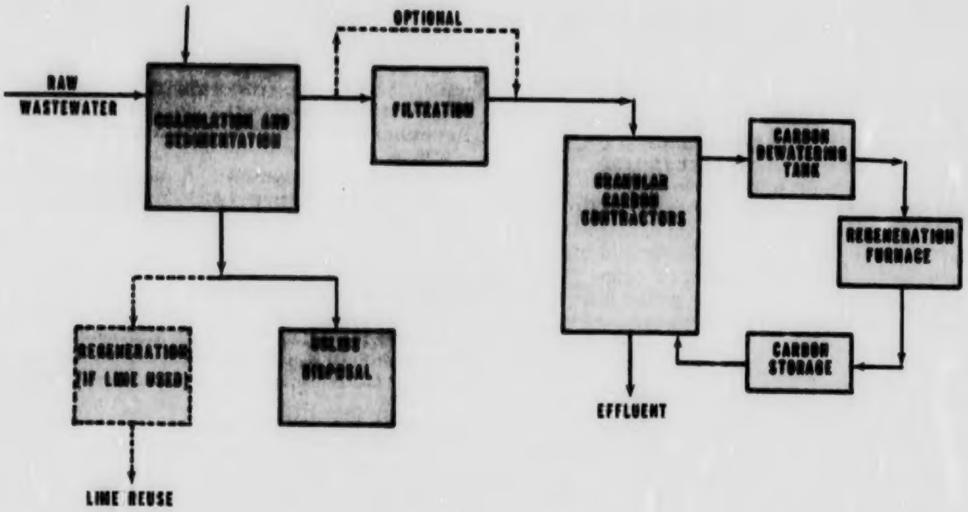

OPTIONAL
RAW WASTEWATER
COAGULATION AND SEDIMENTATION
FILTRATION
GRANULAR CARBON CONTRACTORS
CARBON DEWATERING TANK
REGENERATION FURNACE
CARBON STORAGE
EFFLUENT
REGENERATION (IF LIME USED)
SOLIDS DISPOSAL
LIME REUSE

FIGURE 1
PHYSICAL CHEMICAL ALTERNATIVE TO BIOLOGICAL TREATMENT

TABLE I

ORGANIC REMOVAL

Biodegradable Organics

A Biological Treatment

  Trickling Filter
  Activated Sludge

B Removal of Particulate BOD by Same Process as shown for Suspended Solids

C Activated Carbon Adsorption

Non-Biodegradable Organics and Color

A Activated Carbon Adsorption

B Chemical Oxidation

3 In order for use of activated carbon to be economical, regeneration and reuse is necessary. Granular carbon can be regenerated in a multiple-hearth furnace. Methods for regeneration of powdered carbon are being developed.

C Nitrogen and phosphorus removal occupy major interest because these nutrients tend to be highly soluble, recycle readily, are difficult to remove effectively and are associated with major problems in surface water eutrophication.

TABLE II

NUTRIENT REMOVAL

Nitrogen

A Biological Nitrification - Dentrification

B Ammonia Stripping by Air Contacting

Phosphorus

A Chemical Precipitation with Inorganic Salts

B Biological Removal

C Ion Exchange

Nitrogen can be removed by biological nitrification-denitrification and ammonia stripping by air contacting. Phosphorus can be removed by chemical precipitation with inorganic salts, biological processes, and ion exchange.

Processes are being or will be evaluated at many location . Many pilot-scale and full-scale plants are now in the design and construction stages.

1 Nitrogen can be removed by the progressive biological conversion of nitrogen to nitrogen gas. This involves use of both aerobic and anaerobic biological steps. Nitrification is typically achieved in an activated sludge aeration tank by extending the normal aeration time and by employing lower ratios of BOD to mixed liquor suspended solids than usual. Denitrification results from anaerobic bacteria utilizing oxygen from nitrates and nitrites. Since the nitrified effluent is deficient in carbon, this stage of treatment requires the addition of an organic supplement such as methanol. The denitrification step may be achieved in a separate tank with mechanical mixing or by developing a population of denitrifying organisms within carbon columns or sand filters following the aerators.

2 Ammonia stripping is incorporated into the 2-stage lime precipitation process. Lime is added to the first stage to raise the pH to 11. Effluent from the first-stage clarifier is contacted woth air in a cooling tower for ammonia removal.

3 Phosphorus can be removed by addition of inorganic coagulants at several steps in conventional treatment; raw sewage, mineral addition to the aerator in activated sludge, and by tertiary clarification. Costs for various processes will be discussed later.

4 Chemical coagulation, sedimentation, filtration, and granular carbon adsorption are being evaluated at seven locations at flows up to 7.5 million gallons per day. Plants are located at South Lake Tahoe, California; Piscataway, Maryland; East Chicago, Indiana; Nassau County, New York; Dallas, Texas; Washington, D.C.; and Lebanon, Ohio. The treatment facility at Lake Tahoe has been in operation at 2.5 mgd for several years and has been expanded. Other plants not included are Colorado Springs at 2 mgd and Chicago at 1 mgd.

TABLE III

PILOT PLANT AND FULL-SCALE PLANT STUDIES

CHEMICAL COAGULATION, SEDIMENTATION FILTRATION, AND GRANULAR CARBON ADSORPTION

		FLOW MGD
1	S. Lake Tahoe, CA	7.5
2	Piscataway, MD	5
3	E. Chicago, IN	0.75
4	Nassau Co., NY	0.55
5	Dallas, TX	0.15
6	Washington, DC	0.10
7	Lebanon, OH	0.07

5 A schematic diagram of the Tahoe treatment system is shown on Figure 2. Lime is added in the first stage to raise the pH to 11. This step removes most of the suspended solids, phosphates, and carbonate compounds. Effluent from the first-stage clarifier is then subjected to counter-current air contacting to remove ammonia; effluent from the cooling tower is recarbonated with CO_2 to precipitate excess calcium as $CaCO_3$ in the second-stage clarifier; at this point the pH is dropped to 9.5. Addition of small dosages of alum, polyelectrolyte or activated silica ahead of the filters may be necessary to maintain product clarity in the second-stage effluent. The final treatment step is granular carbon adsorption. Carbon and lime are regenerated for reuse at this facility. Facilities are not provided for demineralization.

6 Figure 3 is a schematic of the Santee, California plant. Nearly all processes being studied for tertiary treatment are represented here. Biological nitrification-denitrification is used with an organic feed source being added

prior to the denitrification step, followed by chemical coagulation, sedimentation and filtration. Flow is 2 mgd. A side stream of 50,000 gal./day will be subjected to additional treatment to produce a water of similar quality to potable water. Treatment includes carbon adsorption, demineralization, and chlorination. Electrodialysis and ion exchange will be studied in parallel for demineralization.

7 Table 4 shows pollutant removal for tertiary processes. Secondary effluent is assumed to have 20 mg/1 suspended solids and 13 BOD. Microscreening or rapid sand filtration will produce an effluent with 6mg/1SS and 8 mg/1 BOD. Lime clarification, filtration and granular carbon adsorption will produce an effluent containing 2 mg/1 of SS and 2 mg/1 of BOD.

TABLE IV

TERTIARY SUSPENDED SOLIDS AND BOD REMOVAL

	Tertiary Effect	
	BOD MG/L	SS MG/L
Rapid Sand Filtration	8	6
Chemical Clarification	6	2
Chemical Clarification and Filtration	5	1
Filtration and Granular Carbon Adsorption	2	2

8 As discussed earlier, phosphorus can be removed by chemical treatment of raw sewage, mineral addition to the aerator in the activated sludge process, and tertiary clarification. Careful study of each design problem is required to determine treatment location and choice of chemicals.

9 Alternate chemical systems are shown here. Iron coagulants are ferrous chloride, commercial or waste pickle liquor, and ferric chloride. Aluminum is added in the form of alum or sodium aluminate. Lime treatment involves use of single-stage or two-stage systems.

TABLE V

CHEMICAL SYSTEMS

A Iron

Ferrous Chloride-Base-W/O Polymer

Commercial or Waste Pickle Liquor	Lime or Sodium Hydroxide
Ferric Chloride - W/O	Polymer

B Aluminum

Alum or Aluminate - W/O Polymer

C Lime - W/O Additives

1 or 2 stage

TABLE VI

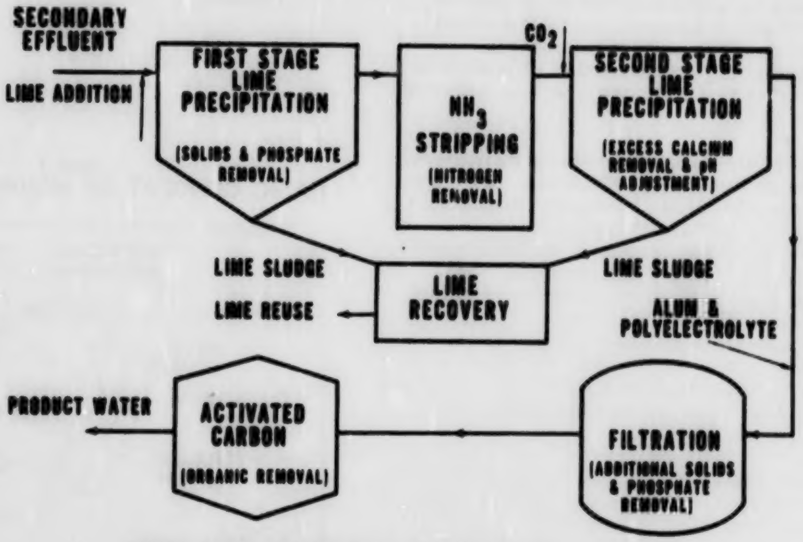
FACTORS AFFECTING CHOICE OF CHEMICALS

1 Influent Phosphorus Level
2 Effluent Discharge Standard
3 Wastewater Characteristics (Alkalinity, etc.)
4 Plant Size
5 Chemical Costs including Transportation
6 Sludge Handling Facilities
7 Ultimate Disposal Alternatives
8 Other Processes Utilized

10 Phosphorus can be removed from raw sewage by the addition of iron, lime, and polymers. Iron salts such as ferrous chloride are used at dosages from 1 to 2.5 mg of iron per mg of soluble phosphate. Chemical costs are about 1.5¢/1000 gal. Total costs are affected by the reduced load to secondary treatment units and will be determined from current studies at three plants. The PEP system or Phosphate Extraction Process marketed by Dorr-Oliver employs lime precipitation in primary tanks.

FIGURE 2

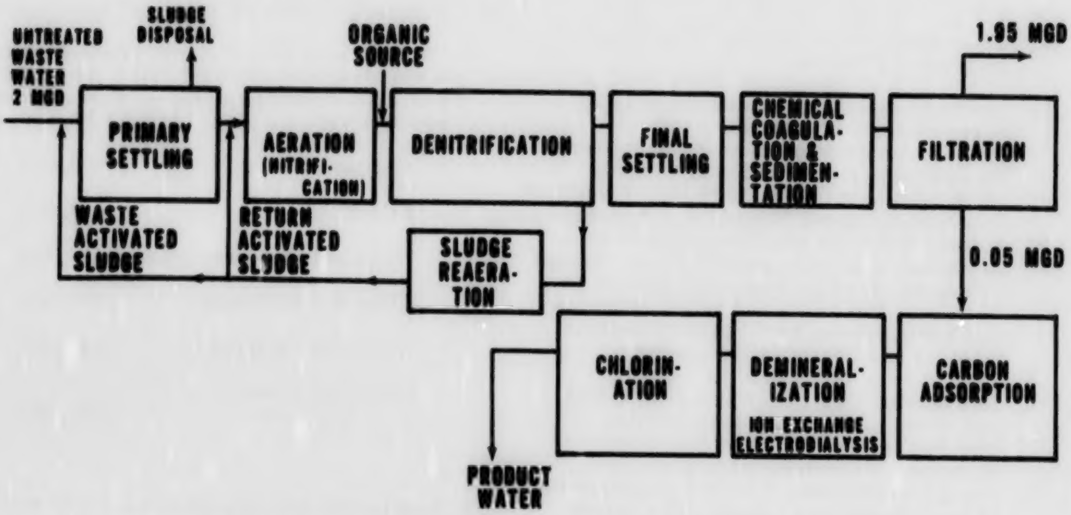
TERTIARY TREATMENT INCLUDING CHEMICAL NUTRIENT REMOVAL


SECONDARY EFFLUENT
LIME ADDITION
FIRST STAGE LIME PRECIPITATION
(SOLIDS & PHOSPHATE REMOVAL)
NH_3 STRIPPING
(NITROGEN REMOVAL)
CO_2
SECOND STAGE LIME PRECIPITATION
(EXCESS CALCIUM REMOVAL & pH ADJUSTMENT)
LIME SLUDGE
LIME RECOVERY
LIME REUSE
LIME SLUDGE
ALUM & POLYELECTROLYTE
PRODUCT WATER
ACTIVATED CARBON
(ORGANIC REMOVAL)
FILTRATION
(ADDITIONAL SOLIDS & PHOSPHATE REMOVAL)

SOUTH TAHOE WATER RECLAMATION PLANT - 7.5 MGD

FIGURE 3

COMPLETE CONVENTIONAL AND TERTIARY TREATMENT


UNTREATED WASTE WATER 2 MGD
SLUDGE DISPOSAL
PRIMARY SETTLING
ORGANIC SOURCE
AERATION (NITRIFI-CATION)
DENITRIFICATION
FINAL SETTLING
CHEMICAL COAGULA-TION & SEDIMEN-TATION
FILTRATION
1.95 MGD
0.05 MGD
WASTE ACTIVATED SLUDGE
RETURN ACTIVATED SLUDGE
SLUDGE REAERA-TION
CHLORIN-ATION
DEMINERAL-IZATION
ION EXCHANGE
ELECTRODIALYSIS
CARBON ADSORPTION
PRODUCT WATER

RESEARCH AND DEVELOPMENT STUDIES AT SANTEE, CALIFORNIA

11 By mineral addition to the aerator, phosphorus concentration in activated sludge process influent can be reduced from 10 to 2 mg/1 or less by adding 1 to 2 parts of aluminum for each part of phosphorus. Chemical costs are about 3¢/1000 gal. for sodium aluminate. Costs including amortization of chemical feeders and maintenance are 4 to 5¢/1000 gal.

12 Results of the full-scale study on mineral addition at Pomona, California. For use of sodium aluminate, overall phosphorus removal on unfiltered samples varied from 78 to 90 percent over an A1:P ratio of 1:1 to 2:1. Slightly better removal was obtained with alum. Ninety percent was obtained at an A1:P ratio of 1.4. A Deterioration of effluent quality in terms of turbidity occurred. The problem was solved by the addition of turbidity occurred. The problem was solved by the addition of 2mg/1 of polymer.

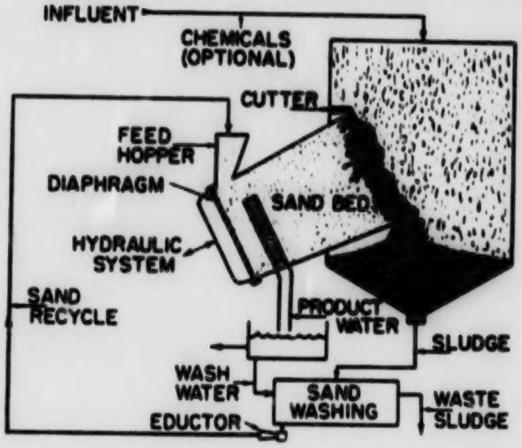
13 Tertiary clarification can be accomplished by use of lime or alum. Overflow rates for lime precipitation are about 1500 gpd/sf for single-stage and 1000 gpd/sf for 2-stage. Lime dosage is dependent on the alkalinity of the water, pH, and the degree of removal required. Single-stage treatment is practical for hard waters at dosages of about 300 mg/1. Soft waters generally require 2-stage treatment at dosages as high as 400 mg/1. Cost of lime at 300 mg/1 is about 1.75¢/1000 gal. Economics of lime recalcination are dependent on disposal methods available.

14 Tertiary clarification with alum is accomplished with an aluminum to phosphorus ratio of about 2. If influent phosphorus concentration is 8 mg/1, alum cost is about 4¢/1000 gal. Overflow rate used is about 700gpd/sf. Approximately twice the clarifier surface area is required as compared to lime. Sludge produced is voluminous and difficult to dewater. Recovery and reuse schemes are being investigated but no practical solution has been divulged to date.

15 The Moving Bed Filter is a new solids and phosphorus removal technique. The Application of this concept for phosphorus removal using alum is being investigated. A schematic diagram of the Moving Bed Filter. The unit is basically a sand filter. As the filter surface becomes clogged, the filter media is moved forward by means of a diaphragm and a clean surface is exposed. The sand and accumulated solids are separated by washing and the sand is reused.

FIGURE 4

BASIC CONCEPT OF MOVING BED FILTER



III COSTS OF ADVANCED WASTE TREATMENT

A Capital costs for advanced waste treatment processes for a 10 mgd plant are as follows: For comparative purposes, cost of a 10mgd activated sludge plant is about $3,600,000.

Microscreening	$ 250,000
Filtration	440,000
2-Stage Lime Precipitation	720,000
Lime Recalcination	640,000
Ammonia Stripping	740,000
Granular Carbon Adsorption	1,600,000

B Total costs for conventional plants and projected costs for some tertiary processes are shown here. Costs are expressed in cents per 1000 gal. and include amortized construction, operation, and maintenance. For a 10 mgd plant, costs for second treatment are 11 cents compared to 7.2 for coagulation - sedimentation-filtration, 8 for granular carbon, and 14 for electrodialysis. Cost of tertiary treatment for removal of nutrients, organics, and inorganics is estimated at 26 to 30 cents/1000 gal. Judged by current standards, costs for tertiary treatment may seem high. However, with the higher degree of treatment, the reclaimed water becomes salable for reuse applications. For water reuse this financial burden would be reduced Further the operations included in physical chemical water renovation are more amenable to manual or automated process control, sludge is more adaptable for dewatering and disposal and equipment space requirements are reduced.

IV IMPROVEMENTS IN CONVENTIONAL PROCESSING

A A full-scale plant study on use of pure oxygen in the activated sludge process has been conducted at Batavia, New York under contract with Linde Division of Union Carbide Corporation. This 2.5 mgd plant has two parallel process trains with separate piping, clarification and return sludge facilities. With recent developments in oxygen production and dissolution technology, oxygen is competitive with air and can be produced economically on site. This scheme is being investigated at additional locations.

1 Shows the oxygen dissolution system demonstrated at Batavia. This is a multi-stage co-current system. Each stage has mechanical mixers and compressors. Note that the system is enclosed in order to maintain high oxygen utilization efficiency.

2 This shows the test facility at Batavia. Temporary covers are installed over the existing aeration tanks. This system has operated at aeration tank detention times of 1 to 1.5 hours with BOD removals of 90 percent or more. Preliminary cost projections indicate a decrease in secondary treatment costs of 20 to 30 percent. Aerobic digestion occurs in the high solids and dissolved oxygen system resulting in an additional cost savings.

3 A study on the effect of flow equalization on performance of the activated sludge process is being conducted at Newark, New York. This new plant has a design flow of 2.8 mgd and an initial flow of about 1.7 mgd. One aeration tank will be utilized for equalizing plant flow with

TABLE VII

POMONA DATA ON MINERAL ADDITION: 2-mgd Plant, April, 1968

Aluminum Form	Ratio Al:P	Overall Phosphorus Removal	
		Unfiltered	Filtered
$NaAl(OH)_4$	1:1	78	85
$NaAl(OH)_4$	1.5:1	85	94
$NaAl(OH)_4$	2:1	90	96
$Al_2(SO_4)_3$	1:1 1.4:1	80 90	83 95

installation of temporary pump and piping (Figure 6). Aeration tank detention time is about 5 hours and will be varied to simulate an overloaded condition by use of temporary bulkheads. Flow equalization has the potential of improving operation of existing overloaded plants as well as new plants. With the advent of advanced waste treatment processes, the economics of flow equalization will become more favorable.

4 Pilot-scale studies are being conducted on application of the U-tube aeration concept to wastewater using aspirated air. The basic U-tube consists of parallel vertical pipes with a return bend. Air is introduced by means of a venturi aspirator or diffuser at the upper end of the inlet pipe. High transfer efficiencies can be obtained at depths from 20 to 40 feet. An alternate design is use of concentric pipes.

a For application in sewer systems, potential plugging problems are reduced by installation downstream from pumping stations. Feasible locations are on the end of long force mains or in the drywell of pumping stations with short force mains. For such systems, headloss through the aspirator and U-tube can be considered in pump design. Compressed air can be utilized in existing systems where sufficient pump head is not available.

b For post aeration of secondary effluent, sufficient head is normally available during periods of low flow in receiving streams when dissolved oxygen levels are more critical.

TABLE VIII

ESTIMATED UNIT COSTS FOR WASTE TREATMENT PROCESSES

		CENTS/1000 GALLON		
CONVENTIONAL PROCESSES	SIZE-MGD	1	10	100
Primary		14	7.5	5.1
Primary & Activated Sludge		20	11	8.6
TERTIARY PROCESSES				
Microscreening		1.8	1.5	1.2
Coagulation & Sedimentation (Lime - Phos. Rem.)		4.0	3.7	3.4
Filtration		7.7	3.5	1.6
Granular Carbon Adsorption		16	8.0	4.0
Electrodialysis		21	14	8.7

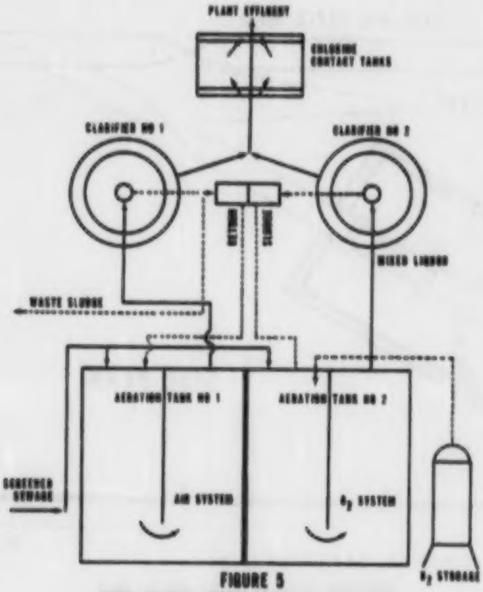

PLANT EFFLUENT
CHLORINE CONTACT TANKS
CLARIFIER NO 1
CLARIFIER NO 2
RETURN SLUDGE
MIXED LIQUOR
WASTE SLUDGE
AERATION TANK NO 1
AERATION TANK NO 2
SCREENED SEWAGE
AIR SYSTEM
O_2 SYSTEM
O_2 STORAGE

FIGURE 5
PURE OXYGEN STUDY BATAVIA, NEW YORK

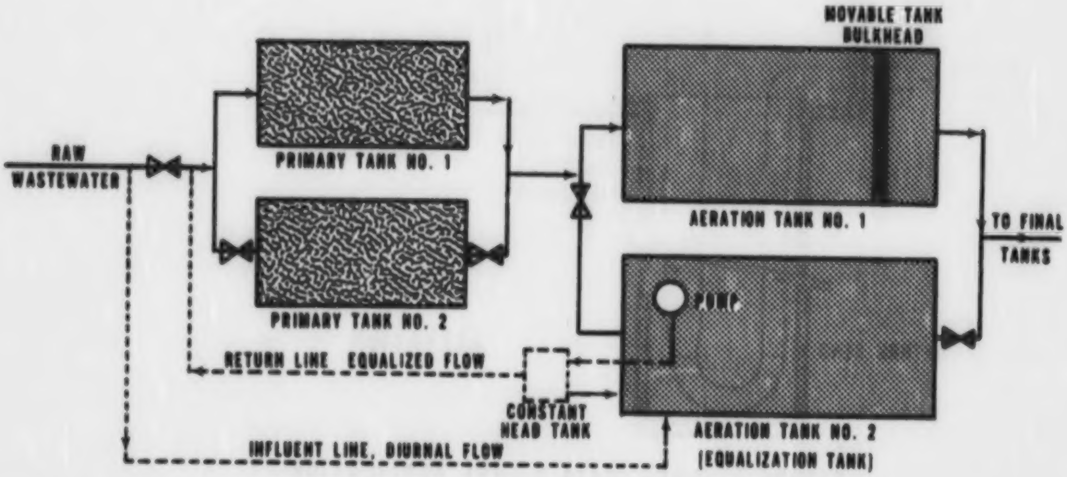

MOVABLE TANK BULKHEAD
RAW WASTEWATER
PRIMARY TANK NO. 1
PRIMARY TANK NO. 2
AERATION TANK NO. 1
TO FINAL TANKS
PUMP
RETURN LINE, EQUALIZED FLOW
CONSTANT HEAD TANK
AERATION TANK NO. 2 (EQUALIZATION TANK)
INFLUENT LINE, DIURNAL FLOW

FIGURE 6
FLOW EQUALIZATION STUDY
NEWARK, NEW YORK

22<

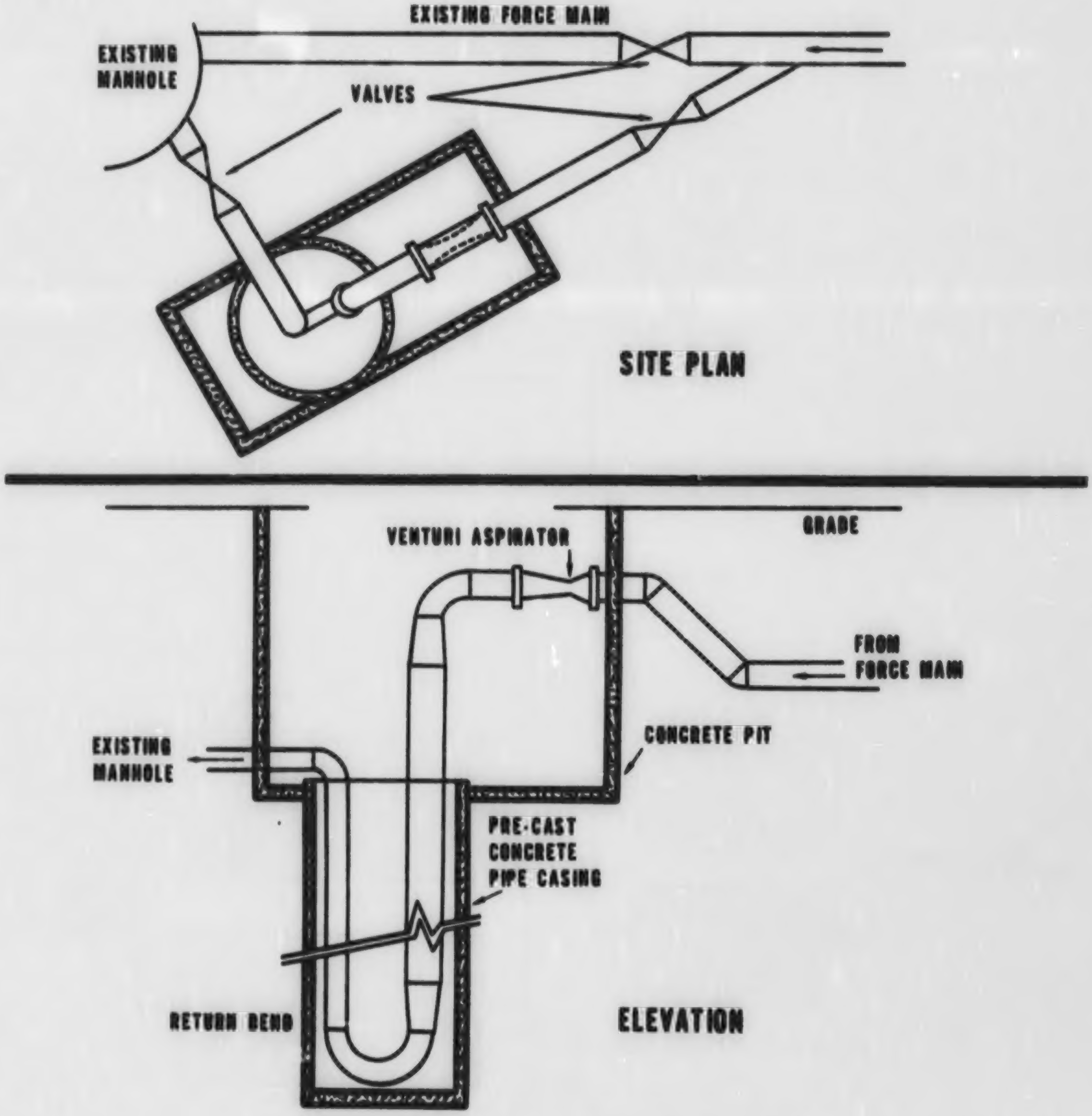

EXISTING FORCE MAIN
EXISTING MANHOLE
VALVES
SITE PLAN
GRADE
VENTURI ASPIRATOR
FROM FORCE MAIN
EXISTING MANHOLE
CONCRETE PIT
PRE-CAST CONCRETE PIPE CASING
RETURN BEND
ELEVATION

FIGURE 7 FORCE MAIN U-TUBE APPLICATION

c Three grant studies on in-sewer treatment for reduction of corrosion and septicity are being conducted. An extensive 3-year study is being conducted by the Los Angeles County Sanitation Districts on the paramenters affecting sulfide generation and oxidation. Included are quantitative evaluations on rate of generation of sulfides in slime layers and the flowing stream, rate of reoxygenation through the air-liquid interface in sewers, and design criteria for aeration of force mains.

5 A study of in-sewer aeration methods is being conducted at Jefferson Parish, Louisiana. Methods to be evaluated include two U-tube aerators on the end of force mains, an eductor or air lift pump in a gravity sewer, and a vortex aerator and venturi aspirator installed in a force main.

a Air and gaseous oxygen are being injected into force mains at Port Arthur, Texas. Use of pressurized aeration tanks at pumping stations is also being evaluated at Port Arthur.

B Mathematical models are or have been developed for conventional and advanced processes. Models developed are sets of computational procedures from which the cost and performance of a process can be computed when certain information is supplied as input. The input supplied is the stream vectors of flow and characteristics of all influent streams to the process and decision parameters which specify the desired mode of operation and basis of design. The program output includes equipment sizes, the magnitude of minor streams such as chemical requirements, and the stream vectors for all effluent streams. Capital cost, operating and maintenance cost, and debt service are also computed.

Models for the activated sludge process were developed at the Advanced Waste Treatment Laboratory in Cincinnati. Research Contracts have been awarded for models for final clarification, coagulation and sedimentation, fluidized bed and multiple hearth incineration, ammonia stripping, electrodialysis, and ion exchange.

V DESIGN MANUALS

A We have recognized the need for improved methods of dissemination of results from research and development programs and are planning a series of design manuals. These manuals will supplement our technical seminars and publications on results of in-house and extramural projects.

The information needed by consulting engineers for design of advanced waste treatment plants is now available for some processes. However, the data are included in numerous publications, reports, and manufacturers' literature. The purpose of the manuals will be to compile available information in a form which can be readily utilized, and provide detailed information on hardware selection and system design.

B Manuals to be developed shown here are phosphorus removal, upgrading capacity and efficiency of existing plants, and treatment of secondary effluent by granular carbon adsorption. The phosphorus manual will include chemical treatment in the primary, mineral addition to the aerator in the activated sludge process, and tertiary clarification. Upgrading existing plants will include suspended solids removal, increase in sedimentation capacity by use of tube clarifiers, and flow equalization. The manual on use of granular activated carbon will be limited to treatment of secondary effluent using pressure or gravity flow.

1 Tentative outline for the design manual is shown in Table 9. It is planned to include sections on basic design considerations, a literature review of research studies and field experiences, outline of laboratory and/or pilot studies needed for design, choice of candidate processes, economic analyses of alternatives, and preliminary plant designs. This section on plant design will include typical plant layouts, conceptual plans and specifications, and special design requirements including corrosion control, valve selection, piping design, control systems, and chemical handling. Personnel requirements for plant operation and maintenance will be included.

TABLE IX

BASIC OUTLINE FOR DESIGN MANUALS

A Basic Design Considerations

B Literature Review - Research Studies and Field Experiences

C Outline of Laboratory and/or Pilot Plant Studies Needed

D Choice of Processes for Further Study

E Economic Analyses of Alternatives

F Preliminary Plant Designs

  1 Plant Layout

  2 Conceptual Plans and Specifications

  3 Special Design Requirements - Corrosion Control, Valve Selection, Piping, Chemical Handling, etc.

G Personnel Requirements

2 Conventional treatment plants have historically been designed in many cases on the basis of flow, BOD, and suspended solids without the benefit of laboratory and pilot plant studies. This approach is not considered feasible for some advanced processes. The manuals will include the necessary lab and pilot plant studies for process selection and design.

3 Design of conventional plants has involved criteria such as lbs. of BOD per 1,000 cu. ft. of aeration tank volume and cu. ft. of air per gallon. A different set of parameters will be required for advanced processes.

TABLE X

DESIGN PARAMETERS

UPGRADING EFFICIENCY OF EXISTING PLANTS

MICROSCREENING

  Choice of Fabric and Configuration

  Washwater Requirements

FILTRATION

  Choice of Upflow, Downflow, Moving Bed, or Gravity Flow

  Media Selection and Depth

  Filtration Rate - GAL/MIN/SQ FT
  Backwash Rate - GAL/MIN/SQ FT

SEDIMENTATION - TUBE CLARIFICATION

FLOW EQUALIZATION

  Basin Sizing

  Mixing and Aeration Requirements

4 Design parameters for granular carbon adsorption systems are shown in Table 11. These include choice of pressure, gravity flow, downflow, upflow, or moving-bed contact systems. Design parameters for systems are carbon particle size, superficial linear velocity, contact time, carbon capacity, and carbon loss on regeneration. Contactor design involves selection of number of stages, and use of concrete or rubber-lined steel vessels. Piping selection and design is more complex because of abrasive materials.

5 Design parameters for microscreening include choice of fabric and configuration, drum speed, and washwater requirements. Filtration parameters include choice of downflow, upflow, moving-bed, and gravity or pressure systems; media selection and depth, filtration rate, and backwash rate. Parameters for flow equalization are basin sizing and mixing and aeration requirements.

6 Design of phosphorus removal systems involves choice of chemical addition point, choice of chemicals, and ultimate disposal alternatives. Hardware selection and design includes control systems, chemical feeding and storage, mixing and flocculation, solids separation, and solids concentration and disposal.

TABLE XI

DESIGN PARAMETERS

GRANULAR ACTIVATED CARBON SYSTEM

Choice of Pressure, Gravity Flow, Downflow, Upflow, or Moving-Bed Systems

Carbon Particle Size

Superficial Linear Velocity - GAL/MIN/SQ FT

Contact Time

Carbon Capacity - LB COD/LB Carbon

Regeneration Loss

Number of Contacting Stages

Contactor Design - Concrete or Rubber Lined Steel Vessels

Piping Selection and Design

VI OTHER ADVANCED WASTE TREATMENT PROJECTS NOW BEING PLANNED INCLUDE:

A Development of Phosphate-free Detergents

B Development and Demonstration of Instrumentation for Automatic Control of Plants

C Additional Studies on Distillation

In summary, costs of tertiary treatment will be high by present standards. Greatly increased expenditures will be required to eliminate pollution of our lakes and streams. Water reuse by industry and for nonpotable purposes will be necessary to reduce the financial burden.

Potable water reuse by mixing with fresh water supplies is considered to be a distinct possibility in the future.

This outline was prepared by C. L. Swanson, Chief, Engineering & Design Branch, Division of Facilities Construction and Operation, OWP, EPA, Washington, DC 20242.

METHODS FOR ACHIEVING HIGH BOD AND SOLIDS REMOVALS FROM CONVENTIONAL PLANTS

I INTRODUCTION

It is an established fact that many of our biological treatment plants produce poor quality effluent. Often, the treatment process itself is mistakenly cited as being inadequate, whereas the true fault lies in not applying known principles and technology efficiently. In other cases, plants perform poorly because they are overstressed out of all proportion to design conditions.

Can we upgrade secondary treatment in general and the activated sludge process in particular? Certainly at many plants we can. In many cases, pollution can be greatly reduced almost overnight by maximum use of existing treatment technology. A properly designed and well-operated activated sludge plant needs no new technology to achieve 90-95% carbon, suspended solids, and coliform removals; what is needed is an awareness of the fundamentals involved and a willingness to pay for and apply all we know.

II PROPER PLANT OPERATION

A Since the inception of secondary treatment, the major emphasis has been on reducing costs, not reducing pollution. For a weakly concentrated process stream such as wastewater, it is difficult to conceive of any scheme that would be more efficient and economical than aerobic microorganisms. However, there is simply no substitution for time and aeration, and we must be willing to pay for the required treatment volume and blower capacity to allow the biological system to perform efficiently. When we skimp on these two necessities, we have cheaper treatment, but also poorer treatment.

B All too often, poor treatment is caused due to the failure of city administrators to realize the importance of providing competent, certified plant operators. A plant will not run itself. The difference between a poor and a good operator can often mean 25-30% difference in overall effectiveness. It really defies explanation to invest huge sums in designing and constructing a plant and then turn it over to a chief operator who has no conception of what is involved in waste treatment and what steps can be taken to meet varying problems. Clearly, local, state, and Federal officials have a responsibility to train, attract, and hold good technicians and operators. This implies, again, that we must be willing to pay a fair salary for value received.

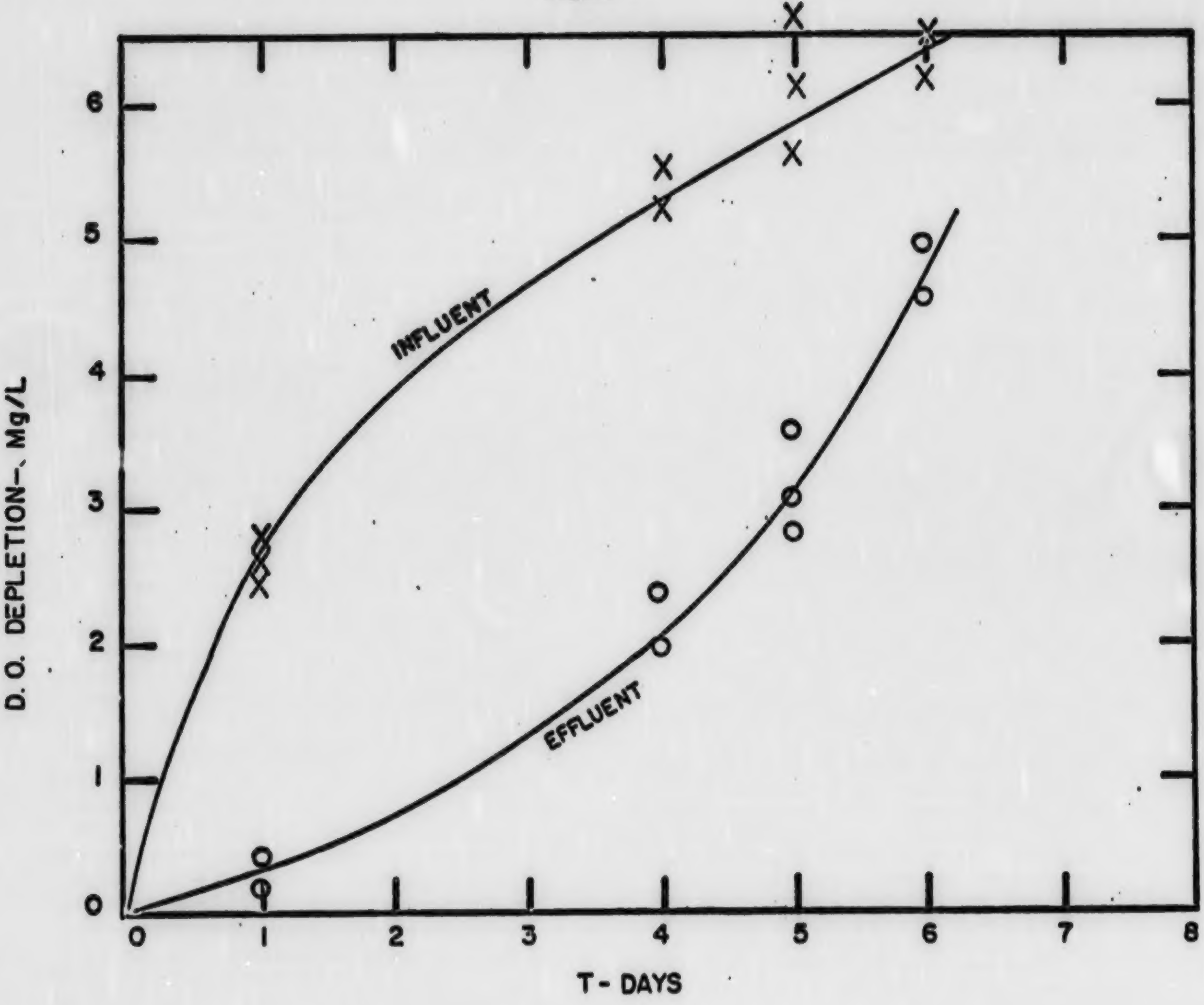
C Once a competent operating staff is assembled, adequate maintenance and replacement budgets must be provided to keep the hardware in good condition. Continuing good operation is not possible when key equipment is constantly down for repairs or becomes antiquated.

D Proper plant operation should include around-the-clock operation and frequent inspection and adjustment, where necessary, of all controllable process flow rates. Many new instruments are on the market which are applicable to sewage treatment. These instruments can assist the operator not only in monitoring his plant, but also, in some cases, automatically controlling certain key parameters such as sludge recycle rates. Magnetic flow meters, variable speed pumps, rate controllers, sludge blanket indicators, and on-stream pH, dissolved oxygen, and temperature detectors and recorders are types of equipment which are being requested in new plant designs.

E And last, but certainly not least, proper operation must include adequate laboratory facilities and staff to make the necessary measurements and analyses to document plant performance. Continuous performance records not only assist in the day-to-day operation and control of the plant, but they are valuable reference information when plant expansion is planned. The day is past, too, when BOD and suspended solids measurements will suffice. The BOD test is strictly a procedure to measure the rate at which oxygen is depleted as organic matter is decomposed. It was never intended to indicate the total quantity of organic matter present in a sample nor to measure treatment efficiency. Yet, we use the procedure in both these contexts today.

1 Figure 1 shows the daily oxygen depletion of an influent and an effluent. One curve is concave and one is conves which means that the reaction rates of the two bear no relation to one another. Comparing the BOD's of these two samples to obtain a measure of plant efficiency is like subtracting apples from oranges.

Figure 1



2 A better estimate of total organic matter and the oxygen demand present is given by COD or TOC, and unoxidized nitrogen. Figure 2 indicates how these tests can be used to predict the ultimate oxygen demand of waste.

3 In addition to carbon, suspended solids, and nitrogen, phosphorus measurements should become routine practice for laboratory staffs. The subject of phosphorus will be dealt with in greater detail later on today. I should note

Figure 2

Ultimate Potential Oxygen Demand of an Effluent

$$UOD = 2.67C + 4.57N$$

UOD = Ultimate Oxygen Demand

C = Organic Carbon Content

N = Total Oxidizable Nitrogen (incl. Organic and Ammoniacal)

Recommended Tests

1. TOC to determine C
2. TKN to determine N

COD may be roughly substituted for 2.67C if TOC Analyzer not available.

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before leaving the subject of analytical procedures that the preceding comments on the BOD test are strictly my opinion.

III AIDING OVERLOADED PLANTS

Let's assume we have a competent operator, a good laboratory staff, and modern equipment and controls, yet overall treatment efficiency is still unsatisfactory. If there are no unusual toxic or grossly atypical wastes present, the impairment can usually be traced to one or more of three factors: hydraulic overload, organic overload, and poor final clarifier liquid-solids separation.

A Hydraulic Overload

Figure 3 shows several techniques which may be tried to combat hydraulic overload in certain situations. They are simple but they work.

1 The most obvious place to start, particularly in small towns, is to locate and reduce needless sources. Industrial cooling water, for example, should never enter the sanitary sewer system. Many industrial rinsing operations can reduce the volume of rinse water if asked. Ground water infiltration is another possible area which may be attacked.

2 Very large interceptors quite possibly can serve as temporary holding tanks during the day and the backup processed through the plant during low nighttime flows.

3 Perhaps the most direct and efficient way is to install a surge or equalization tank ahead of the plant. The flow to the plant can be smoothed considerably. If an outpumping scheme is used as shown in Figure 4, the flow can be kept virtually constant over a given 24-hour period. It is well to provide some minimal aeration to keep the waste from going septic. In certain instances by utilizing a floating takeoff and outpumping, a primary settler can also serve as a surge tank, though generally they are smaller than needed.

When all else fails, additional capacity must be provided.

B Organic Overload

Possible remedies for this problem are given in Figure 5.

1 A surge tank is useful in diluting strong wastes and evening out the organic loading to the secondary system.

2 Industries in smaller towns are usually willing to program their more noxious

Figure 3

Control of Hydraulic Overload

---

1. Find and reduce needless sources
2. Consider using large interceptors as holding tanks
3. Construct surge tank
    - Fancy
    - Plain
4. Enlarge plant

Figure 4

Surge Tank Sizing

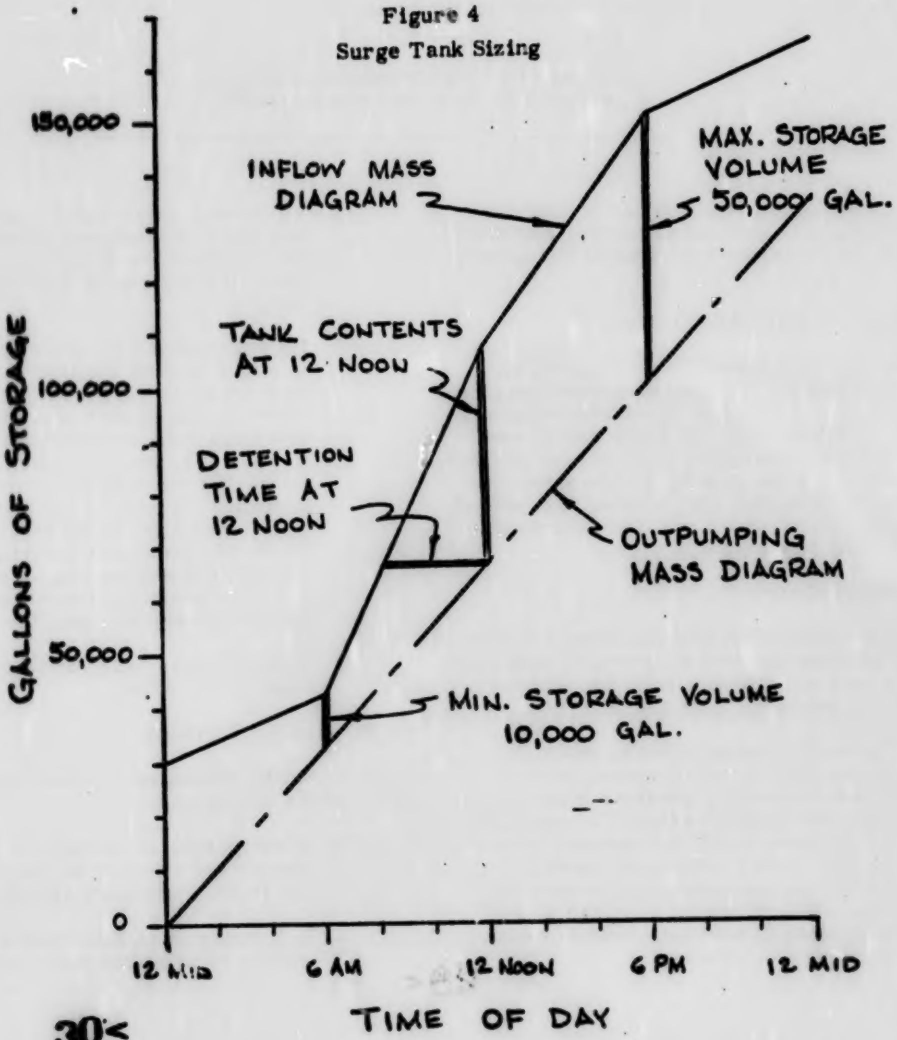

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Figure 5

Control of Organic Overload

1. Surge tank to smooth load
2. Have industry program load
3. Program digester supernatant recycle or eliminate
4. Install roughing filter-in line or loop operation
5. Consider two-stage treatment

waste discharges to coincide with lower loading periods of the day.

3 Probably the most frequent and upsetting shock load plants have to deal with is digester supernatant. COD and ammonia concentrations of 10,000 - 20,000 and 1,000 mg/l, respectively, are common and together constitute a tremendously high oxygen demand. If the supernatant recycle cannot be broken by separate treatment, it should be returned slowly to the main plant flow stream, not in a slug in a one-hour period in the middle of the day as is often done.

4 The recently developed plastic media have a good potential for taking out the bulk of incoming organics when used as a roughing filter ahead of the normal secondary process. If used strictly to control organic overloads, a roughing filter does not require a separate settler.

5 Two-stage treatment is worth considering when a high rate system is to be enlarged. The two sludge systems will enhance the possibility of achieving nitrification, and it is unlikely that both sludges will be adversely affected by toxic spills.

C Liquid-Solids Separation

1 Physical features that will improve liquid-solids separation are suction-type sludge removal mechanisms, variable speed pumps tied to a sludge blanket indicator or the incoming raw flow meter signal, scum removal equipment, and proper baffling to prevent short circuiting and strong velocity currents.

2 Polymers can temporarily keep a bulking sludge from "washing out" gross mixed liquor solids. A drum of polymer and a small dosing pump, if kept in reserve, can get a plant through an upsetting shock load of several days duration. Polymers are stable for long periods of time and are available in powder and liquid form.

IV A RECENT DEVELOPMENT FOR UPGRADING OVERLOADED PLANTS

Union Carbide Corporation has recently completed a research contract with FWPCA which successfully substituted oxygen for air in the activated sludge process. A closed system is employed to achieve high oxygen utilization. Dissolved oxygen is readily maintained in the system even with mixed liquor suspended solids concentrations up to 8,000 mg/l. The necessary aeration volume to achieve efficient organic synthesis and oxidation is reduced to 2-3 hours. Figure 6 shows a schematic diagram of the system and Figures 7 and 8 give treatment data at two nominal aeration periods of 3 and 1-1/2 hours, respectively, which were studied during the contract. This process has very good potential for increasing the capacity and improving efficiency of many overloaded plants in this country without enlarging aeration tank volumes.

V SUMMARY

A summary of anticipated effluent quality and cost data for conventional processes is given in Figure 9. Note that as effluent quality improves, the cost associated with the process rises. Process 6 as shown i.e., primary

Figure 6

SCHEMATIC DIAGRAM OF UNION CARBIDE OXYGENATION SYSTEM

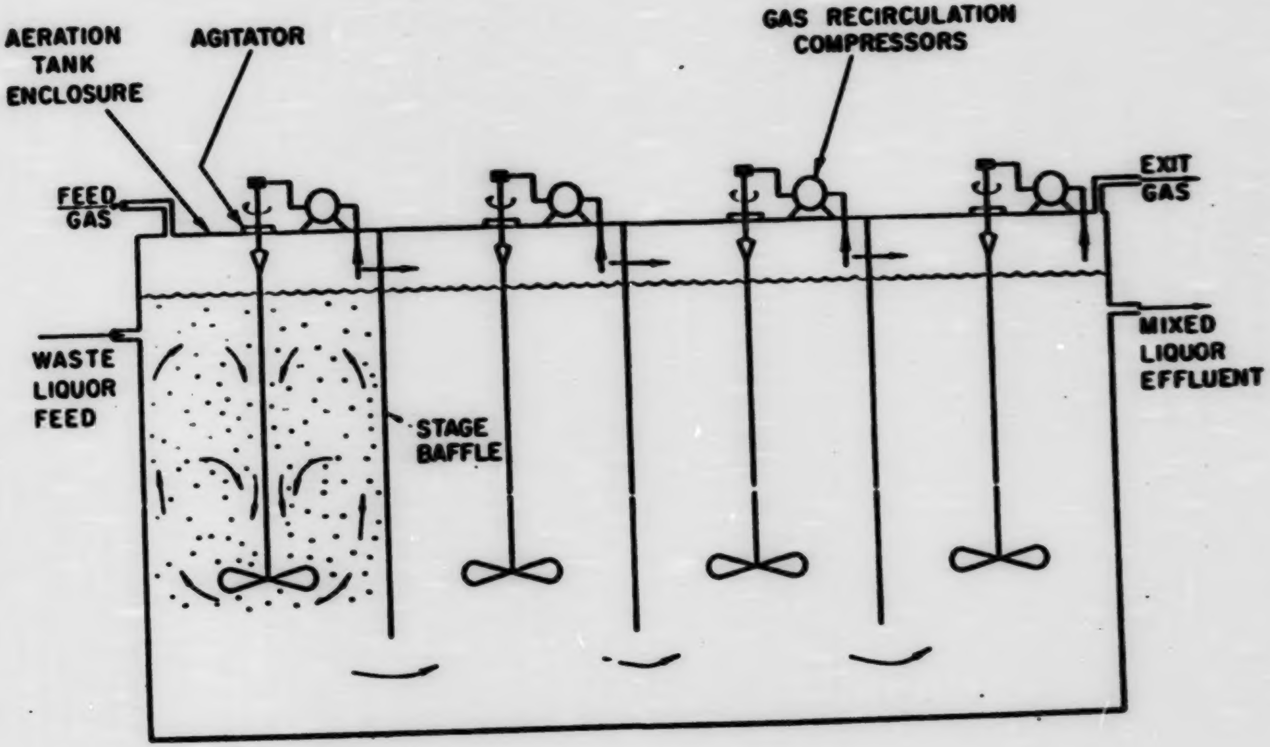

AERATION TANK ENCLOSURE
AGITATOR
GAS RECIRCULATION COMPRESSORS
FEED GAS
EXIT GAS
WASTE LIQUOR FEED
STAGE BAFFLE
MIXED LIQUOR EFFLUENT

Figure 7

PHASE NO. 3 OPERATION
SUMMARY OF DAILY AVERAGE PERFORMANCE (PRELIMINARY)

Parameter	Air Train	Oxygen Train
Raw Sewage Feed Rate (Mgd)	1.30	1.30
Return Sludge Recycle Ratio	0.26	0.45
Raw Sewage Temperature (°F)	69	69
MLSS Concentration (ppm)	3400	6800
MLVSS Concentration (ppm)	2410	4375
Recycle Sludge TSS (ppm)	15400	22050
Mixed Liquor D.O. Concentration (ppm)	0.9	8.6
Clarifier Effluent D.O. Concentration (ppm)	0.0	6.9
TSS of Clarifier Effluent (ppm)	31	16
% TSS Removal	92.5	96
BOD of Clarifier Effluent (ppm)	42	19
% BOD Removal	85	93
COD of Clarifier Effluent (ppm)	147	69
% COD Removal	72	87
Ft^3 Air/Gal. Raw Sewage	3.91	-
Lbs. Oxygen Utilized/Lb. BOD Consumed	-	1.0
% of Feed Oxygen Utilized	-	93.5
Lbs. Dry Solids Wasted	2900	650
Total Oxygenation System Power (HP)	-	18.3
Nominal Aeration Time (Hrs.)	3	3

23<

Figure 8

PHASE NO. 2 OPERATION
SUMMARY OF DAILY AVERAGE PERFORMANCE

Parameter		Oxygen Train
Raw Sewage Feed Rate (mgd)		2.62
Return Sludge Recycle Ratio		0.33
Aeration Detention Time (Hrs.)	1.5 Nominal	1.17 Actual
Raw Sewage Temperature (^oF)		66
MLSS Concentration (ppm)		7142
MLVSS Concentration (ppm)		4575
Recycle Sludge TSS (ppm)		29600
Sludge Volume Index		35.0
Mixed Liquor D.O. Concentration (ppm)		9.8
Clarifier Effluent D.O. Concentration (ppm)		1.0
TSS of Clarifier Effluent (ppm)		18
% TSS Removal		90.6
BOD of Clarifier Effluent (ppm)		22.9
% BOD Removal		90.1
COD of Clarifier Effluent (ppm)		99.0
% COD Removal		70.0
% Phosphate Removal		22.2
Lbs. Oxygen Utilized/Lb. BOD Consumed		1.03
% of Feed Oxygen Utilized		92.6
Lbs. Dry Solids Wasted		1804
Total Oxygenation System Power (HP)		16.8

Figure 9

EFFLUENT CHARACTERISTICS AND COST DATA FOR CONVENTIONAL PROCESSES

Process	Total operating cost: $/mil. gal. 10 mgd plant	Expected Ranges of Effluent Characteristics in mg/l					
		BOD_5	COD	TOC	SS	TKN	TP
1. PS	60-65	115-150	200-275	65-90	75-100	25-35	8-12
2. PS + TF (high rate)	85-90	40-60	100-150	30-50	30-50	20-30	7-11
3. PS + TF (low rate)	90-95	20-35	55-90	20-30	20-35	15-20	8-12
4. PS + AS (high rate)	95-105	30-40	75-100	25-35	30-35	20-30	7-11
5. PS + AS (conventional)	110-115	15-25	40-70	15-25	20-25	15-20	7-11
6. PS + AS (operated to nitrify)	120-130	10-20	30-60	10-20	15-20	2-4	8-12

Legend

PS - Primary sedimentation
TF - Trickling filtration
AS - Activated sludge
SS - Suspended solids
TKN- Total Kjeldahl nitrogen as N
TP - Total phosphorus as P

Assumed Raw Wastewater Quality

BOD_5 : 175-225 mg/l
COD : 325-450 "
TOC : 100-130 "
SS : 200-250 "
TKN : 30-40 "
TP : 10-15 "

settling followed by activated sludge operated to nitrify, represents about the maximum efficiency that can be extracted from biological treatment alone.

This outline was prepared by R. C. Brenner, Research Sanitary Engineer, Biological Treatment Research Program, Advanced Waste Treatment Research Laboratory, Robert A. Taft Water Research Center, NERC, EPA, Cincinnati, OH 45268.

35<

STATUS OF ACTIVATED SLUDGE PROCESS

I CURRENT STRESS

A Increasing emphasis on secondary treatment and up-grading effluent quality has resulted in many new plants, additions to existing plants and conversion of other processes to activated sludge.

B There has been a long history of operational problems connected with the activated sludge process.

1 Design often stressed tankage and economy rather than performance of a dynamic biological system. Versatility and workability during operation often were reduced by inadequate understanding of process and operational realities or subsequent "cuts" in proposed facilities.

2 The lack of knowledge of operating personnel in energy and growth characteristics and operating control techniques often results in low grade effluent quality. Public apathy, inadequate operational budgets for manpower, materials, and maintenance often relegate the facility to a "monument" status instead of a functional entity.

C Population density and activity changes have forced a reevaluation of the "known". Design changes necessarily must follow to up-grade environmental quality.

D Many alternates are available for the activated sludge process. The versatility of the process and possibilities for supplementation for effluent quality improvement with respect to general or specific components cause it to be ranked very high among these alternates. The process is not dead yet; many of its potentials are not generally recognized.

II THEORY OF ACTIVATED SLUDGE

A The process was first reported in 1914 by Ardern & Lockett (1).

B Chemical vs biological theories caused much discussion.

1 Butterfield in 1935 (2) isolated Z. ramigera from activated sludge and developed pure culture activated sludge from Z. ramigera (3).

2 McKinney isolated several floc forming bacteria in 1950 and demonstrated that flocculation was associated with the stage of development (4).

C Energy concept for bacterial flocculation.

1 Work at the Massachusetts Institute of Technology (MIT) in 1951 developed relationships between food (energy) levels, growth and flocculation (4).

2 During a situation involving high food concentrations (energy), cell growth was rapid but flocculation was nil.

3 After available food had been metabolized into a lower free energy situation, flocculation occurred.

4 Protozoa assisted clarification.

D Recent research

1 The Z. ramigera concept has been supported.

2 The accumulation of poly beta hydroxybutyric acid (PHB) in the cell has been associated with flocculation.

3 Flocculation has been observed without PHB formation. Naturally occurring polyelectrolytes and other factors are associated.

E Basic Concepts

1 Bacteria metabolise organics with synthesis of new cells.

a Simple, soluble organics are metabolized first.

b Colloidal and suspended organics metabolized more slowly.

c A portion of the organics are oxidized to yield energy for synthesis.

d There is no metabolism without synthesis.

e A definite quantitative relationship exists between energy and synthesis.

2 At low energy levels, agitation promotes floc formation.

3 Endogenous metabolism of cell mass and storage products provides energy for regrowth and oxidation thus decreasing sludge mass.

4 Protozoa feed on bacteria and further reduce sludge mass; also improves clarification by pickup of scattered residues.

5 Overall reaction is synthesis followed by endogenous respiration.

6 All biochemical reactions are both quantitative and reproducible.

III MATHEMATICAL CONCEPTS AND DESIGN

A In the 1950's, efforts were directed toward mathematical design concepts (5).

B Eckenfelder's Equations

1 Initially developed in 1954

a BOD loading factor $= \frac{\text{lbs BOD/day}}{\text{lbs MLSS}}$

where: BOD = BOD_5 day

MLSS = mixed liquor suspended solids in process.

For calculation (90% removal)

$= \text{BOD}_5 \times 1.5$ taken for La

$$= \frac{24\,La}{Sa\,t\,(1 + r)}$$

where: La = influent BOD-Average 24 hrs/day
Sa = average MLSS
t = aeration time (hrs) based on (Q + R) = (1 + r)
r = sludge recycle ratio
Q = inflow volume
R = recycle volume

b Lbs O_2/day = aLr + bMLVSS

where: a = 0.5 = lbs MLVSS produced/lb BOD removed
b = 0.08 = endogenous rate of sludge oxid./day
Lr = BOD removed

c Lbs volatile solids produced/day = 0.55Lr - 0.08MLVSS

d Oxygen uptake rate (mg/1/hr) $= \frac{\text{lbs } O_2\text{/day}}{24\,Va}$

where: Va = aeration tank capacity in million lbs water

2 1970 Concepts

a BOD loading factor $= \frac{F}{M} = \frac{24So}{Xat}$

where: F = food
M = microorganism mass
So = influent BOD in lbs/day
Xa = average MLVSS
t = aeration time in hrs.

b Substrate removal rate coefficient

$$k = \frac{So - Se}{XatSe}$$

where: Se = effluent BOD

c Lbs O_2/day = $a'Sr(Q/V) - b'Xa$

$= QKi + aSrQ - (bXaV + QXe)$
(oxygen basis)

where: a' = fraction BOD oxidized
b' = fraction VSS oxidized

d Lbs sludge produced/day

where: Q = waste flow
Xi = influent SS
Xe = effluent SS
a = fraction BOD converted to cells
b = fraction VSS oxidized (dry solids basis)

e Simultaneous equations are used to estimate unknowns. Experimental data may require alteration of certain equation components. It is usually advisable to design for the longer of the aeration time estimates.

Determination of constants

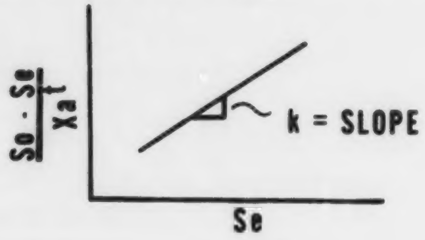


FIGURE 1

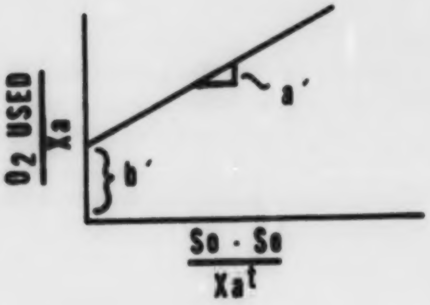


FIGURE 2

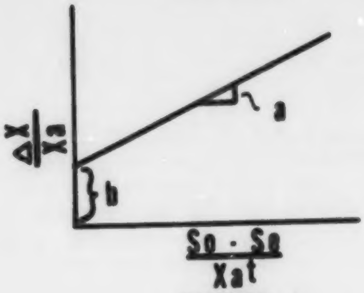


FIGURE 3

where: data are obtained from variable loadings

C Modified Monod Equations

1 In 1942, Monad developed set of empirical equations to approximate pure culture bacteria growth.

2 In 1950, Monad expanded his equations to continuous feed systems.

a Mass growth

$$\ln(X/Xo) = (\mu - D)t$$

where: X = microbial mass at time t
Xo = initial microbial mass
μ = specific growth rate
D = dilution rate
t = reaction time

b Specific growth rate (μ)

$$\mu = \mu_o \frac{S}{S_k + S}$$

where: μ = specific growth rate at substrate conc. S

μ_o = maximum specific growth rate

S = substrate concentration

S_k = substrate concentration when $\mu = \mu_o/2$

c Substrate utilization

$$\frac{dS}{dt} = D(So - S) - \frac{X}{R}\left(\mu_o \frac{S}{S_k + S}\right)$$

where: R = microbial mass yield constant

3 Popular with University researchers

4 Problems arose in determination of μ_o, R, and S_k and in failure to consider endogenous respiration.

5 Reynolds and Yang in 1966 modified Monod's equations for activated sludge

a Required MLSS (X)

$$X = \frac{Y(So - S)}{1 + \alpha - \alpha\left(\frac{1 + \alpha}{\alpha + w}\right) + Ket}$$

where: Y = sludge yield coefficient

α = sludge recycle ratio

w = waste sludge ratio (fraction of raw waste)

Ke = endogenous respiration rate

b Desired aeration time (t)

$$t = \frac{1}{XKe}\left\{Y(So - S) - X\left(1 + \alpha - \alpha\left(\frac{1 + \alpha}{\alpha + w}\right)\right)\right\}$$

c Oxygen uptake rate

$$\frac{dO}{dt} = (1 - 1.42Y)\,\frac{So - S}{t} + 1.42\ KeX$$

d Determination of Ke, Y, u_o, and S_k

D Complete Mixing Equations

1 First applied to activated sludge by Garrett & Sawyer in 1952 laboratory evaluation (6).

2 The first detailed set of design equations for complete mix activated sludge (CMAS) plants, was published in 1962 (7) and modified in 1969 (8).

a Organic reduction

$$F = \frac{Fi}{Kmt + 1}$$

where: F = unmetabolized BOD (available BOD at any given time)

Fi = influent BOD

Km = metabolism factor (15/hr @ 20°C)

t = aeration time in hrs

b Active microbial mass

$$MA = \frac{KsF}{\frac{1}{ts} + Ke}$$

where Ma = active microbial mass (mg/1 MLVSS)

Ks = synthesis factor (10.4/hr @ 20°C)

ts = sludge retention time in hrs.

ke = endogenous respiration factor (0.02/hr @ 20°C

c Endogenous mass

Me = 0.2KeMats

where: Me = endogenous mass (mg/1 MLVSS)

d Inert Organic mass

Mi = Inf. Mi(ts/t)

where: Mi - inert organic mass (mg/1 MLVSS)

Inf. Mi = influent inert organic suspended solids (mg/1 VSS)

(domestic sewage: 0.4 VSS in influent

Experience: .20-.40 20 years ago
.4 works better now

e **Inert Inorganic mass**

$$Mii = Inf.\ Mii\ (ts/t) + 0.1\ (Ma + Me)$$

where: Mii = inorganic MLSS (mg/1)
Inf. Mii = influent inorganic suspended solids (mg/1)

f **Total MLSS**

$$M_T = Ma + Me + Mi + Mii$$

g **Oxygen Uptake rate**

$$\frac{dO}{dt} = \frac{1.5(Fi - F)}{t} - \frac{1.43(Ma + Me)}{ts}$$

where: $-\frac{dO}{dt}$ = oxygen uptake rate (mg/1/hr)

h **Temperature effect**

Km, Ks, and Ke change by factor of 2 with 10°C change (over a temp range of approximately 10-35°C)

i **Determination of Km, Ks, and Ke** laboratory and field data

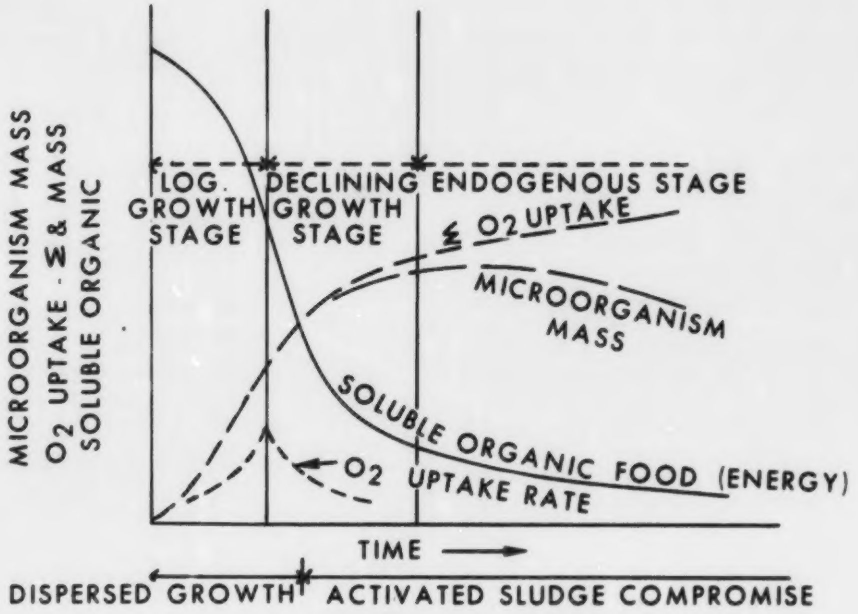
E **Application of design equations**

1 There is considerable confusion as to application to real systems.
2 All 3 sets of equations can yield reasonable design criteria. How the equations are used is more important than which one.
3 Eckenfelder and Monod equations empirical and require separate evaluation for each waste.
4 CMAS equations developed from fundamental microbial concepts

  a Declining growth phase only (food limiting)
  b Single set of design factors for all wastes through evaluation of Ma, Me, Mi, and Mii
  c Variations in design factors related to types of microbes

    1 Bacteria-protozoa yield optimum results
    2 Fungi yield more cell synthesis and less oxidation
    3 Proper environmental design insures proper microbes

FIGURE 6



METABOLISM KINETICS

IV PLUG FLOW VS COMPLETE MIXING (9)

A Longitudinal mixing

1 Conventional activated sludge uses long, narrow aeration tanks with spiral flow aeration to minimize longitudinal mixing.

2 Horizontal velocity approx. 5 fpm along the tank axis.

3 Normally, length to width ratio in excess of 5:1.

4 Axial dispersion concepts used in 1963 to evaluate mixing.

5 Existing studies indicate that current activated sludge tanks are neither plug flow nor completely mixed in spite of a rash of reports that existing aeration tanks are completely mixed.

B Complete mixing

1 Developed in Bury, England, 1919, then built plants long and narrow.

2 First real use was with extended aeration systems in 1950.

3 Fundamental microbial research at MIT in 1950's indicated that CMAS was preferred from a microbial point of view.

4 Laboratory and field studies during past 20 years confirm this.

5 Field CMAS plants have produced results not attainable with any other biotreatment system.

a Alkaline cotton textile wastes, pH 10-11, treated without chemical preneutralization and minimal nutrient additions.

b Phenolic wastes at 2,000-3,000 mg/1 reduced to 0.04 mg/1 in single stage.

c Antibiotic wastes at 7,000 mg BOD/1 reduced to under 50 mg/1.

d Shock loads handled with no significant impairment in effluent quality up to 2-3 times average BOD load.

6 Any desired effluent can be produced from any influent concentration.

7 Not readily accepted by engineers initially for fear of short circuiting.

8 Computer studies by University professors on CMAS vs plug flow indicated CMAS not as efficient as plug flow.

9 CMAS slowly being accepted by engineers.

10 Largest CMAS plant currently under construction is 100 MGD plant at Rochester, New York.

11 CMAS is not instantaneous mixing, arbitrarily defined as having a one minute mixing time at high loading.

12 Fluid flow pattern.

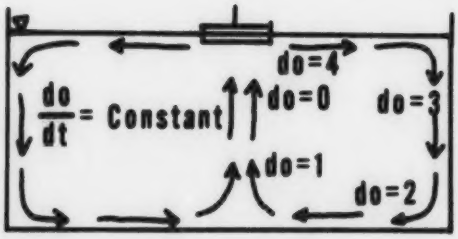


FIGURE 7

SS measurements not accurate enough for use

DO varies along the circulation pattern; lowest on return to the aerator.

Oxygen uptake rate is constant for uniform loading.

13 CMAS most efficient and most economical biotreatment system currently in use and can meet any reasonable effluent BOD requirement.

V AERATION SYSTEMS

A Turbine aeration

1 Diffused air added under turbine.

2 Diffused air establishes fluid flow pattern, not the turbine.

3 Quirk in 1962 found optimum results when air blower power equaled turbine power.

4 Transfer 2.5 to 3.0 lbs 0_2/HP/hr with 20% efficiency.

5 A turbine requires 50% more power when pumping liquid than when pumping a less dense mixture of liquid and gas.

6 Surface aerator above the turbine is not an efficient design.

B Surface aeration

1 General acceptance of surface aeration in the United States has been slow. Diffused air systems were popular and early attempts at surface aeration were penalized by large safety factor regulations. Both designs of units and concepts have improved.

2 The surface aerator essentially is a fluid pump designed to create a large gas-liquid interface area and fluid mixing energy.

3 Maximum oxygen transfer efficiency occurs when the pumped fluid is saturated with oxygen as it leaves the aeration device and returns to it with a large oxygen deficit. Microbial viability constraints require that returning fluid $DO > 0$.

4 Oxygen transfer relationships:

$$\frac{DO}{dt} = K_{La} \, (0_S - 0_L)$$

where: $\frac{DO}{dt}$ = the change in DO per unit time.

K_{La} = The oxygen transfer coefficient
= A measure of turnover time for fluid to be saturated with oxygen.
= Q_s/V under conditions where Q_s or quantity of fluid flow saturated with oxygen
= Q_w or quantity of fluid flow pumped. (maximum transfer).

V = Pump flow rate.
Fluid turnover time = V/Q_w

O_S = the oxygen saturation value for the fluid under test.

O_L = the DO value of the liquid returning to the aeration device.

5 Draft tubes and baffles help to direct flow properly and improve mixing.

a A draft tube orients upward fluid flow to the surface aerator, maintains solid suspension and minimizes dead spots of low DO fluid at the bottom of the tank.

b Baffles tend to break up centrifugal momentum of tank contents that may produce short circuiting.

6 Surface aerator performance depends upon the nature of the liquid, the unit, its size, loading, and tank geometry. The oxygen transfer capabilities at maximum loading are most important. For example: With similar devices.

a 3.0 to 4.0 lbs 0_2/HP/hr may be transferred in pure water.

b 2.5 to 3.0 lbs O_2/HP/hr may be transferred in mixed liquor.

c 1.5 to 2.0 lbs O_2/HP/hr may be transferred in large aerated lagoons.

C Rotor aeration

1 Developed in Holland as modification of Kessener brush.

2 Current design: 28" dia. rotor with 12 rows of 2x6" flat tabs @4" centers.

3 At 70 RPM with 3 to 6" submergence 2.5-4.0 lbs O_2/HP-hr may be transferred.

4 A modification at the University of Kansas resulted in a 60" dia. rotor which can transfer between 4.0 and 5.0 lbs O_2/HP-hr.

5 These units are more specialized than other aerators with primary use in shallow tanks.

D Pure oxygen

1 Increased driving force and higher saturation level are the main factor.

2 Early studies by Okun not economically feasible.

3 Studies at the University of Kansas in the 1960's demonstrated value of pure oxygen lay with partial pressure and oxygen demand. Without a high oxygen demand rate, it was not economical.

4 Recent interest revived by Union Carbide study at Batavia, N.Y.

5 Contrary to Batavia reports, there is no fundamental basis for lower sludge production in pure oxygen in contrast to air provided adequate air is supplied. With limited air, endogenous respiration seems to be suppressed and metabolism of adsorbed suspended solids are retarded, both resulting in higher than normal sludge production.

6 A properly designed CMAS plant with mechanical surface aeration unit is more economical than pure oxygen at present time both in capital and operating costs.

VI HIGH MLSS SYSTEMS

A Recent trend is toward higher MLSS concentrations, 4,000 - 6,000 mg/l.

B Extended aeration plants with 5,000 - 10,000 mg/l MLSS started trend in effort to reduce frequency of sludge wasting.

C Eckenfelder's equations indicated a MLSS-time relationship. Aeration time could be shortened with increased MLSS for the same effluent quality.

D CMAS equations indicated short aeration times would require high MLSS concentrations but effluent quality would decrease.

E High MLSS concentrations require sludge return with weak organic wastes but not for concentrated wastes. Domestic sewage and normal activated sludge require 100% sludge return rate based on raw waste flow for 5,000 mg/l MLSS and 200% for 6700 mg/l.

F Inert solids in raw waste determine inert solids in mixed liquor. Both inert organic and inert inorganic suspended solids build up in mixed liquor. Domestic sewage normally contains 125 mg/l of inert suspended solids. Primary sedimentation will remove this 125 mg/l down to 50 mg/l while chemical coagulation can reduce it to less than 10 mg/l.

G High MLSS concentration systems require special sedimentation tanks to prevent excessive solids loss in effluent.

H Current design should limit MLSS to lowest level based on microbial mass for peak organic loads, 2,000 - 4,000 mg/l MLSS. High solids are likely to require higher aeration cost than are saved by reduction of sludge solids for disposal.

VII OXIDATION DITCH

A Developed by Paasveer in Holland for small extended aeration systems.

1 Utilize a rotor aerator for agitation and creation of air-liquid interface area.

2 Employs a shallow oval ditch as a fluid container.

3 Depth of ditch limited by fluid pumpage and minimum velocity to keep sludge in suspension, (1.0 fps).

4 Distance between rotors determined by oxygen demand rate and time of flow.

B The oxidation ditch has become popular for animal feedlot waste treatment.

1 Hogs and beef cattle are grown on slotted floor directly above the ditch.

2 Concentrated urine and manure drop directly into ditch along entire length for relatively uniform feeding.

3 Hog manure has BOD of 40,000 mg/l which must be diluted by treated wastes in ditch if the system is to remain aerobic.

4 Rotors spaced at intervals depending upon the total manure load and time of travel between rotors.

5 High nitrogen content of manure and urine, long aeration periods, and excess DO produces nitrification.

6 As organic load increases, oxygen demand exceeds available DO and denitrification results with loss of nitrogen. At full organic load, it is possible to obtain essentially complete nitrification and denitrification.

7 The oxidation ditch potential is just beginning to be realized.

VII SOLIDS SEPARATION

A The effluent quality of activated sludge systems are determined by effluent suspended solids.

B The present trend is to reduce surface overflow rate to prevent suspended solids loss but with a constant retention period, the depth is reduced.

C With high MLSS, the trend to shallower final sedimentation tanks is an invitation to disaster.

D Concern is centered on the rapid removal of activated sludge from the sedimentation tank.

1 Hydraulic sludge collection devices have been developed, but return sludge concentrations are difficult to control.

2 Units operate best at 6,000 mg/l suspended solids or less.

3 They require high return sludge rates for proper MLSS in aeration.

4 They require separate sludge thickening system for waste activated sludge.

5 High sludge return rates result in mixed liquor displacement from the aeration tank creating turbulence in the sedimentation tank.

6 Turbulence is a definite problem in shallow units.

7 Sludge retention time in final sedimentation tank actually is the same as the corresponding fluid retention time.

8 No real advantage in hydraulic sludge return system over properly designed unit.

E Conventional sedimentation tanks.

1 Normal concentration in gravity unit and sludge scrapers is 10,000 mg/l.

2 Higher sludge concentrations than 10,000 to 12,000 mg/l indicates excessive storage in sedimentation tank and definite sludge deterioration or higher than normal fraction of inert inorganics.

3 Minimum MLSS concentration results in minimum return sludge and minimum turbulence at inlet to the sedimentation tank.

4 With HIGH MLSS, the sedimentation tank will be at least half full of sludge at all times regardless of sludge collection mechanisms.

5 Sludge retention time the same as corresponding fluid retention time.

6 The volume of sludge in the sedimentation tank points up need for deep tanks.

7 All final sedimentation tanks should be equipped with scum removal equipment and scum baffles around wiers.

F Waste activated sludge should be handled separately from primary sludge.

1 Mixing with primary sludge adversely affects primary sludge dewatering.

2 Waste activated sludge is difficult to concentrate or dewater directly as a result of the structure of the microbial cells and hydration of them.

3 Large quantities of chemicals needed to overcome organic characteristics of activated sludge.

VIII AEROBIC SLUDGE DIGESTION

A Waste activated sludge aerobic digestion.

1 Used in extended aeration and package contact stabilization plants. Part of the digestion has occurred in process.

2 Only active mass can be digested up to maximum of about 80%.

3 Stabilization rate: 2%/hr at 20°C, 1% @ 10°C and 4% @ 30°C.

4 A series of compartments in digestion tank produces plug flow and maximum sludge digestion.

B Primary sludge

1 Reactions are the same as in regular activated sludge aeration tank including rapid metabolism followed by endogenous respiration.

2 Similar to animal manure treatment.

3 Need dilution below oxygen transfer limitations.

4 After 80-90% reduction in active mass sludge, the fluid concentrates by gravity to above 5% and dewaters rapidly.

IX CURRENT DESIGN CRITERIA

A Domestic Sewage

1 Normal flow and BOD variations favor CMAS systems.

2 Any of the 3 sets of mathematical equations will work satisfactorily.

3 Major problem is sludge handling. Every effort should be made to minimize sludge production, reducing sludge disposal to a minimum.

4 Design of any waste treatment system is not complete until all solids are returned to land and fluid returned to the water environment.

5 Trend to short aeration time results in excessive sludge production.

6 With domestic sewage, 3 hours appears to be normal lower limit for aeration time to produce satisfactory effluent at all times and absorb peak loads.

7 Reducing the aeration tank detention below 3 hours is false economy as increased operational costs and equipment costs will override the small savings in aeration tank volume.

8 Oxygen equipment must be designed to meet peak oxygen demand in mixed liquor and not average oxygen demand in pure water. Valid performance specifications are essential for aeration equipment.

9 Phosphate removal can be accomplished best by a combined chemical biological system employing coagulant ahead of primary sedimentation tank to remove maximum quantity of primary sludge prior to aeration. Activated sludge removes residual phosphate with chemical ppt as final polisher if needed.

10 Reduction in aeration costs and sludge handling costs will offset chemical cost for phosphate reduction.

11 Ammonia stripping and low BOD effluents are not compatible.

12 Complete nitrification requires long sludge retention time, over 5 days, with 50% excess oxygen and adequate buffer to keep pH over 6.5.

13 Mechanical surface aeration is the most economical system at present.

14 Aeration systems must be designed to handle peak loads based on actual oxygen uptake rates.

15 Final sedimentation tanks can produce good results at 800 gal/sq.ft./day surface overflow rate with minimum depth of 12 ft.

16 Variable sludge return capability between 25 and 100% based on raw waste flow are essential. Adequate measuring devices also are necessary for control.

17 Sludge wasting should be continuous where possible from the return sludge stream.

18 Waste activated sludge should not be returned to primary sedimentation tanks or anaerobic digesters if at all possible.

19 Aerobic sludge digestion is the best volume reduction method with 15-20 day aeration required, including the sludge retention time in the activated sludge aeration tank.

B Industrial Wastes

1 No fundamental difference in design for domestic sewage or industrial wastes. In both cases, the system must be designed on the basis of the available food energy in process, the oxygen uptake rate exerted and the amount of working biological solids. Total suspended solids carried in process depend upon inert residues, time energy relations, and effluent quality sought.

2 The pH of raw waste important only as it affects pH of mixed liquor. Keep pH between 6.5 and 9.5 in the mixed liquor.

3 Surge tanks help level load variations and produce a better effluent.

4 CMAS systems are best able to handle shock loads that occur from time to time.

X CURRENT OPERATIONAL CRITERIA

A Without good operation, the best design is of little value.

B The current trend toward complex, automated, highly instrumented controls with infinite variability are undesirable. Design should minimize mechanical equipment.

C Mechanical equipment should be selected for ease of maintenance, reliability, long life, and ease of replacement.

D DO control unnecessary but time clock adjustment at 8 to 12 hour intervals can assist in economical control.

E Constant rate of sludge return simplifies operation and produces best results.

F CMAS aeration tanks produce uniform oxygen demand within limits and permit use of constant speed mechanical aerators.

G Solids level control in sedimentation tank and balance between BOD load, sludge wasting, and oxygen utilized permits easy operational evaluation.

XI RECOMMENDED DESIGN FOR DOMESTIC SEWAGE

A CMAS system.

B 3 hrs aeration based on raw waste flow.

C Constant speed mechanical aerator.

D Constant rate of sludge return.

E Constant rate of sludge waste.

F Return sludge to land and water to environment.

G Simplicity of design and operation can produce an effluent with 20 mg/l BOD and 20 mg/l SS without difficulty, 10 mg/l BOD and 10 mg/l SS with a little effort, and 5 mg/l BOD and 5 mg/l SS with close control of design and operation.

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STATUS OF OXIDATION POND PROCESSES

I DEFINITION

The term "Oxidation Pond" or "Waste Pond" may be applied to any quiescent body of water into which wastes are intentionally introduced. The terms, however, are usually reserved for impoundments of waste water which are created by construction of earthwork embankments and excavations or combinations thereof into which wastes are conveyed by pressure conduits, gravity sewers, or open canals.

II POND TERMINOLOGY AND APPLICATIONS

A Applied terms tend to be associated with applications; terminology is loose, often misapplied, and subject to change according to performance and neighborhood response. Lack of understanding, vagaries of the situation, changes in loading, or neglect contribute to the situation. Ponds are common where land costs are lower. Low operational costs are attractive, but the essentials must not be neglected; resulting unfortunate developments lead to derisive terms that tend to be remembered. The following discussion includes some of the items contributing to terms and applications.

B The relationships between dissolved molecular oxygen and microbial activity provide a logical basis for pond classification.

1 Aerobic - contains dissolved molecular oxygen thruout the water mass at all times.

2 Anaerobic - Designed and operated for use in the absence of detectable molecular oxygen in the water mass.

3 Facultative - The most common type of pond is characterized by a dissolved oxygen excess in upper layers of water and a molecular oxygen deficiency in lower layers of the pond.

C Overflow characteristics provide another basis of terminology according to usage:

1 Terminal ponds have no surface overflow. The terminal pond may be further classified as percolating when the water loss mainly occurs by movement into the substrata or evaporative if that happens to be the main reason for prevention of out flow. Both are usually involved.

2 Overflowing or detention ponds have an inlet and outlet with some form of basin between.

3 Intermittent ponds may have a surface overflow during wet weather or high inflow rates,with no overflow at other times.

D Common terminology may be associated with inflow characteristics or sources such as:

1 Sewage ponds or lagoons
2 Feed lot or agricultural drainage
3 Canning, paper mill, meat packing or other industrial drainage ponds.

E Ponds may be classified according to their stage of use in an integrated treatment system.

1 An equilization pond may be used to smooth the flow for subsequent operations.

2 A settling pond may be used to decrease gross solids discharges into other operations and/or to decrease organic loading.

3 The pond may follow primary, secondary or advanced treatment. A polishing pond may be used to back up any treatment operation to improve regular operation or decrease effects of contingencies.

4 A series of ponds may be designated in terms of relative member in the series, such as primary, secondary, tertiary, quaternary, etc.

5 The pond may be used for holding organic sludge,storm water, silt or process sludges, and named accordingly. The pond may be a terminal disposal site to be abandoned when filled, or an holding pond subject to subsequent reprocessing and takeout.

F Loading rate may be the basis of another system of terminology:

1 A high-rate pond may have a high rate of input relative to volume.

2 Low-rate ponds usually have low volumetric, or organic loading relative to volume capacity or time.

3 Intermediate rates may be anything between.

4 The rate concept may be based upon detention time, volume, area, depth, organic content, food/microorganism ratios, solids content, etc.

G The predominant microbiologied activity occurring, or of interest, provides a basis for pond classification:

1 A stabilization pond covers almost any pond class involving mixed growth, detention time and degrees of treatment.

2 An algal pond includes those in which algae can and do grow. This usually means a detention time of 3 or more days and conditions suitable for growth. A long detention time may permit deposition and clarification. Dispersed growth in the outflow characterizes intermediate detention times.

3 A reduction or treatment pond may be characterized by high activities of reducer organisms, such as bacteria fungi or slime organisms. Growth or circulation may lead to diminished light penetration limiting algal activity. These may be considered contact, holding, or high rate ponds. They may be predominantly anaerobic or aerated to maintain circulation and molecular oxygen content. If sludge is returned, or liquid recycled, a new terminology opens.

H Ponds may be associated with one or more large variety of purposes that lead to applied terms:

1 If it is associated with open space, green belt, or community planning, the actual use may change terminology drastically.

2 Wild life propagation, wet lands, marsh development are another basis for terms and use.

3 Ponds may be used for water, land, nutrient, energy recycle, or as a barrier for salt water intrusion. Since ponds eventually fill, land reclamation for park, residential or industrial activities may become progressive uses with water renovation and recharge during early events.

4 Industrial ponds may be used for cooling, recycle water, fire protection, surface irrigation, solids reclamation, pumped storage, or treatment of source or wastewaters. Retention and control of storm water may be intended.

III FUNCTIONAL MECHANISMS AND DESIGN

Waste component stabilization, solids disposal and liquid retention are fundamental considerations in pond design and use.

A Rigorous evaluation of the hydraulic regime including inflow, outflow, percolation and evaporation are first considerations in capacity requirements.

1 The general relationship for ponds (or other wastewater treatment) is:

$$\mu A = Q \qquad (1)$$

where μ is a loading or dissipation velocity
Q is the volume of liquid entering or leaving the pond.
A is pond area

2 The instantaneous velocity of all water entering a pond is $\mu_{in} + i + P_i$

where $\mu_{in} = Q_{in}/A$
i = the rate of precipitation (rainfall)
P_i = rate of groundwater infiltration.

3 The instantaneous velocity of water leaving the pond is $\mu_{out} + P_o + E$

where: $\mu_{out} = Q_{out}/A$
P_o = the rate of exfiltration or percolation
E = the evaporation rate

4 The hydraulic balance for any pond at any instant is:

$$\mu_{in} + i + P_i = \mu_{out} + P_o + E \qquad (2)$$

or regrouped:

$$\mu_{in} - \mu_o = (E - i) + (P_o - P_i) \qquad (3)$$

5 The volume of water leaving the pond by overflow is a major concern because of its influence on the receiving water.

$$\mu_{out} = \mu_{in} - [(E-i) + (P_o - P_i)]$$

a For an overflowing pond:

$$\mu_{in} > (E-i) + (P_o - P_i)$$

b For a non overflowing pond

$$\mu_{in} < (E-i) + (P_o - P_i)$$

B Non overflowing ponds

1 To have a ponding system which never has a surface overflow, it is essential to have a variable input or a variable surface area to provide for storage.

2 Net evaporation $(E-i)$ or E_n) generally is available from long term records of precipitation and evaporation.

Evaluation of month by month data as in Table 1 gives an estimate of E_n.

a From Fig. 1 the net evaporation per year is 44.5 in. and in July reaches a maximum of 130 inches. The mean rate is 0.122 in/day.

b 0.122 in/day represents 3320 gal/A/day.

c At a per capita use of 80 gal/cap/day in the situation in Table 1, evaporation alone will dispose of the wastes of 41 persons per acre throughout the year. During July, the wastewaters of 121 persons per acre could be evaporated.

3 As indicated in Figure 1, a pond receiving water at the rate of 0.1 in./day would have to be about 38 in. deep to avoid becoming dry before the next rainy season and could increase in depth by 25 in. during the rainy season. For a minimum depth of 3 ft., as recommended, the design depth of slightly over 6 ft. is indicated by E_n data.

TABLE I

TYPICAL ANALYSIS EVAPORATION-PRECIPITATION

Month	Evap. E inches	Precip. i inches	E-i inches	(E-i) Year 1	Cumulative Year 2
Jan.	1.1	5.1	-4.0	-4.3	+40.0
Feb.	2.0	4.3	-2.3	-6.6	+37.7
Mar.	3.6	3.6	-0.0	-6.6	+37.7
Apr.	5.6	2.1	+3.5	-3.1	+41.2
May	8.2	0.5	+7.7	+4.6	+48.9
June	9.0	0.0	+9.0	+13.6	+57.0
July	10.8	0.0	+10.8	+24.4	+67.8
Aug.	9.8	0.0	+9.8	+34.2	+77.6
Sept.	7.8	0.0	+7.8	+41.0	+85.4
Oct.	5.0	0.3	+4.7	+45.7	+90.1
Nov.	2.3	1.4	+0.8	+46.5	+90.9
Dec.	1.2	4.1	-2.9	+44.0	+87.2

Figure 1 shows these data in graphic form.

Figure 1.

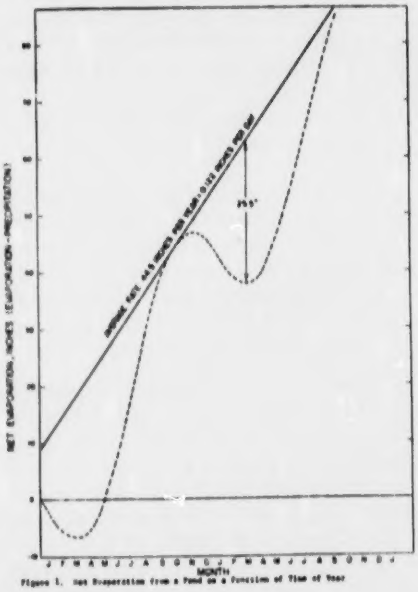


Figure 1. Net Evaporation from a Pond as a Function of Time of Year

4 Net percolation rate (P_o-P_1) may be estimated reasonably well on operating ponds but is difficult to estimate for design purposes.

a If ponds are located above groundwater level, intrusion is unlikely; P_1 becomes zero and P_o or water moving toward the ground water level depends on soil permeability, pressure, area and boundary behavior. Divert storm inflow.

b Table II includes observed percolation plus evaporation rates for a number of operating ponds in California and various soil characteristics.

c Percolation rates may also be estimated by shutting off inflow and outflow or diversion for a 24 hour period without rainfall. The depth measurements before and after indicate percolation plus evaporation if the water level is above the groundwater table. Pan evaporation measurements simultaneously permit evaporation loss estimation (pan evaporation rates commonly are about 1.25 pond evaporation). The net percolation rate is available by difference.

d Careful metering of inflow and outflow permit percolation rate estimates as in c without flow interruption if evaporation rate and surface levels are determined.

e Percolation tests as described for septic tanks (1) do not extend over enough time to be of value in pond design. A test developed, used, and still under evaluation by the author for the past 10 years is outlined in Figure 2. This test is more rigorous because it includes simulation of hydraulic head, continual submergence for 14 days, bottom lining and conditions close to the actual situation. A plot of loss per day

TABLE II

NET PERCOLATION PLUS NET EVAPORATION RATES IN VARIOUS PONDS AND SOIL TYPES IN CENTRAL CALIFORNIA

POND SYSTEM	POND SIZE ACRES	POND DEPTH INCHES	SOIL TYPE	APPROXIMATE TERMINAL LOSS RATE*	
				INCHES DAY^{-1}	GALLONS ACRE^{-1} DAY^{-1}
Woodland South (Ave.)	2	60	Heavy Silty Clay	0.40	11000
American Canyon	4		Coastal Silt		
July-Aug.				0.46	12500
Feb.-Mar.				0.25	6800
Esparto	2	72	Light Silty Clay		
July-Aug.				0.66(1)	18000
Feb.-Mar.				0.25	6800
St. Helena	19	144	Greyish Loam		
Woodland North (Ave.)	2	60	Alkaline Silt	0.73	20000
Woodland Pilot (Ave.)	1/4	60	Alkaline Silt	0.75	20200
Gonzales (Ave.)	3	72	Fine Sand	1.30	35000
Rio Lindo (Ave.)	1/2	72	Gravel with Silt	1.50	41000
Berryessa Resorts (Summer)	1/4	36-72	Course Shale	2.5	68000

*Includes net evaporation which may average 0.1 inch per day and equal 0.3 inch per day during summer. Percolation rates apparently decrease in winter and increase in summer and seem to be influenced by viscosity and gas production in the bottom.

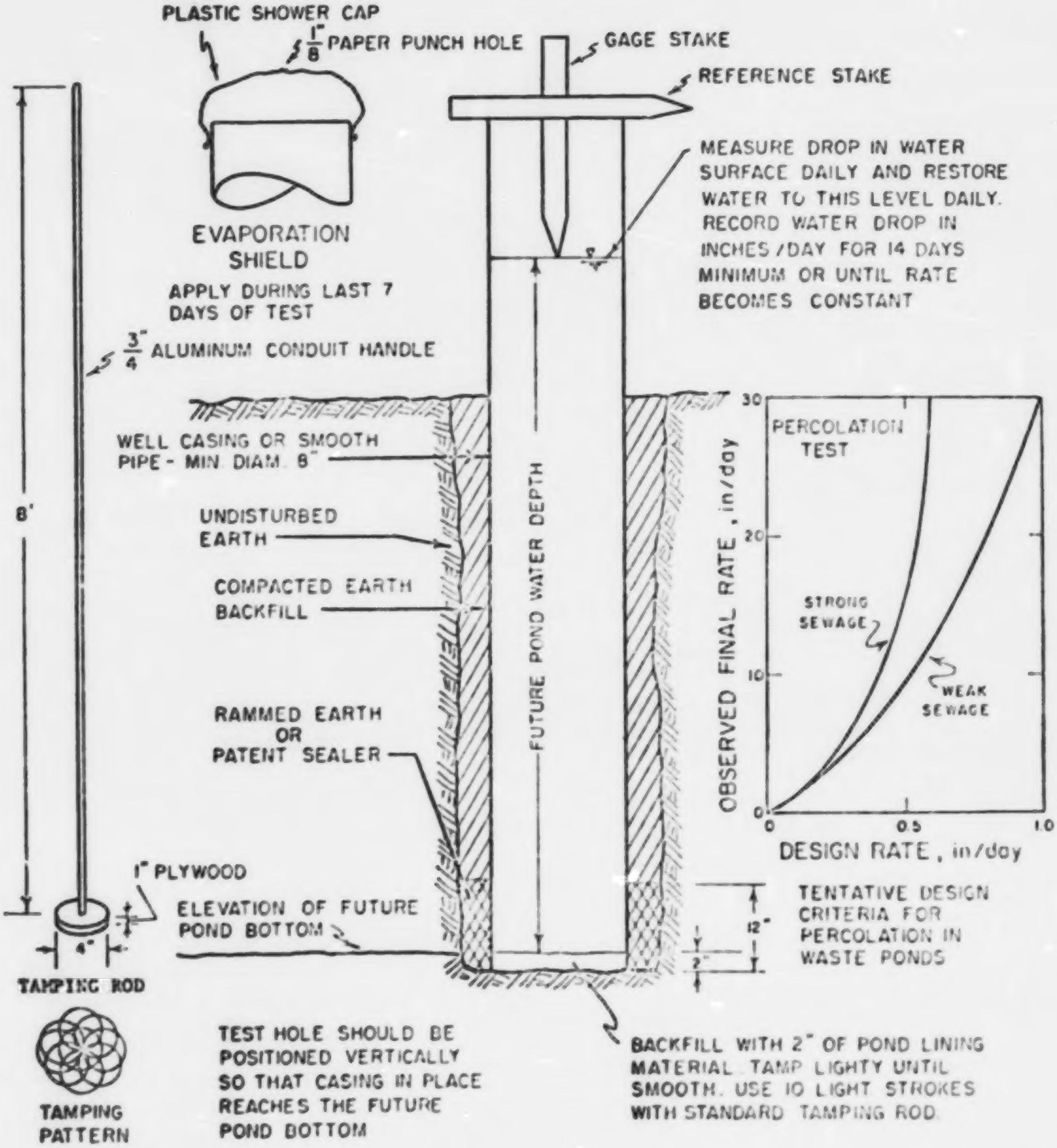
(1) July-August evap. = 0.35 in per day.

(2) In Central California evaporation and precipitation are usually approximately equal in March and hence, net evaporation is 0. Therefore, March values are mainly, Wet Percolation P_o.

may indicate the approach to long time equilibrium if conditions remain uniform. Experience has shown that data obtained from the test as in Figure 2 is valid to be used on strong sewage percolation rates on a long term basis. Comparative tests and pond performance seem to yield conservative data.

f Currently available hydraulic loss rates in central California ponds (Table 2) generally approach 0.5 in + or - 0.25 inches with some exceptions. A loss rate of 0.5 inches per day represents 13600 gal/A/day or the wastewater load from 136 persons/A/day at 100 gal/cap/day water use rate.

PERCOLATION TEST FOR WASTE STABILIZATION POND LOCATION EVALUATION AND FOR ESTIMATE OF GROUND DISPOSAL RATES


PLASTIC SHOWER CAP
$\frac{1}{8}$" PAPER PUNCH HOLE
EVAPORATION SHIELD
GAGE STAKE
REFERENCE STAKE
$\frac{3}{4}$" ALUMINUM CONDUIT HANDLE
8'
WELL CASING OR SMOOTH PIPE - MIN. DIAM. 8"
UNDISTURBED EARTH
COMPACTED EARTH BACKFILL
RAMMED EARTH OR PATENT SEALER
FUTURE POND WATER DEPTH
1" PLYWOOD
4"
ELEVATION OF FUTURE POND BOTTOM
TAMPING ROD
TAMPING PATTERN
12"
2"
PERCOLATION TEST
OBSERVED FINAL RATE, in/day
DESIGN RATE, in/day
30
20
10
0
0
0.5
1.0
STRONG SEWAGE
WEAK SEWAGE

APPLY DURING LAST 7 DAYS OF TEST

MEASURE DROP IN WATER SURFACE DAILY AND RESTORE WATER TO THIS LEVEL DAILY. RECORD WATER DROP IN INCHES/DAY FOR 14 DAYS MINIMUM OR UNTIL RATE BECOMES CONSTANT

TENTATIVE DESIGN CRITERIA FOR PERCOLATION IN WASTE PONDS

TEST HOLE SHOULD BE POSITIONED VERTICALLY SO THAT CASING IN PLACE REACHES THE FUTURE POND BOTTOM

BACKFILL WITH 2" OF POND LINING MATERIAL. TAMP LIGHTY UNTIL SMOOTH. USE 10 LIGHT STROKES WITH STANDARD TAMPING ROD

FIGURE 2 PERCOLATION TEST SET-UP FOR WASTE POND DESIGN

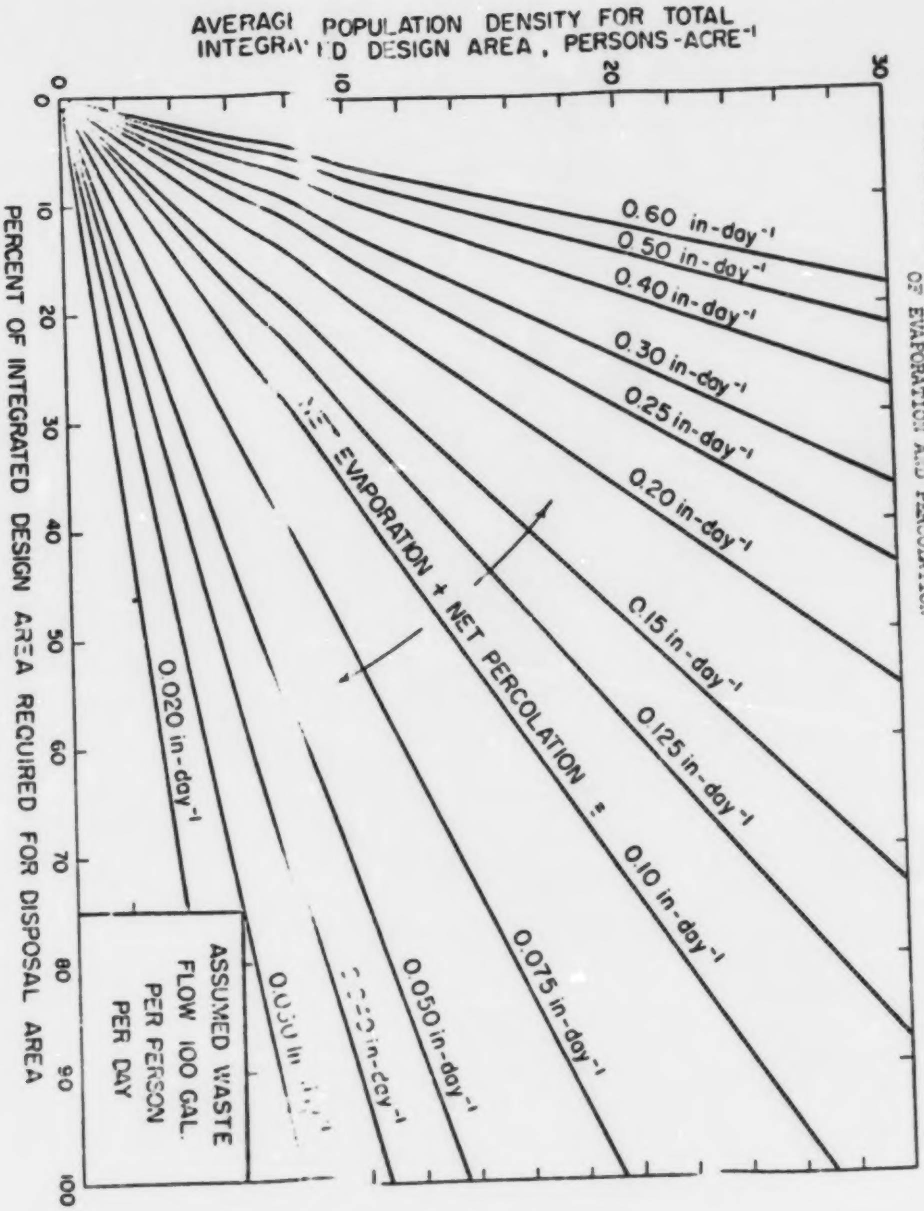
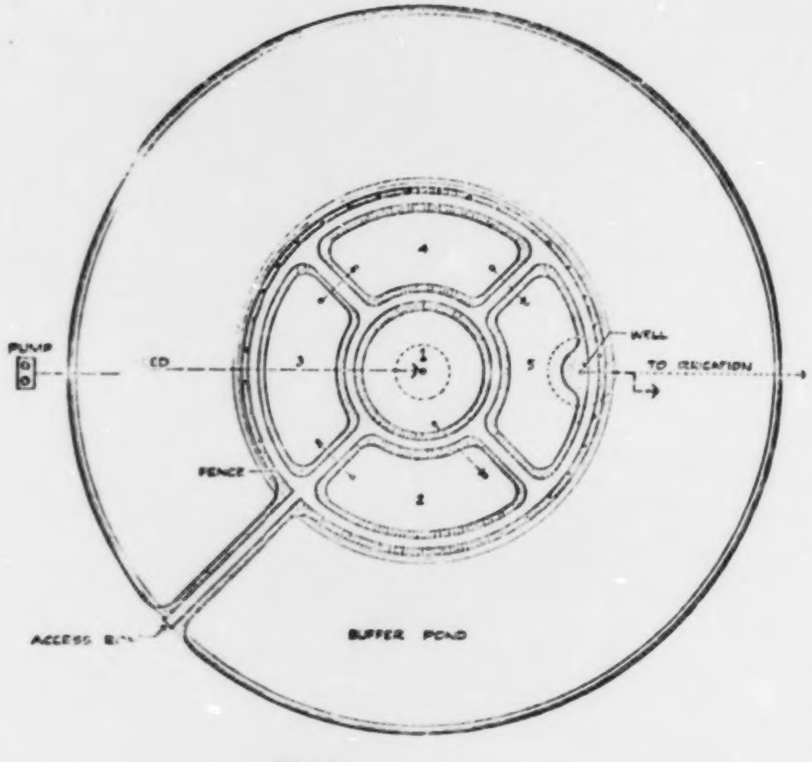


FIGURE 3. RELATIONSHIP BETWEEN POND AREA AND POPULATION DENSITY FOR VARIOUS RATES OF EVAPORATION AND PERCOLATION

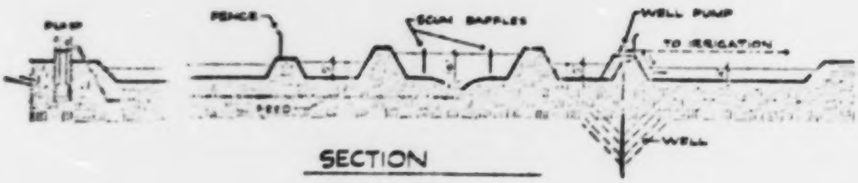
54<

C Pond - Park integrated systems for total disposal.

1 No overflow ponds includes total disposal without affecting surface receiving water in the vicinity. The author (2) and an amplification by Noren (3) describes integrated systems based upon the data in Figure 3 where population density and hydraulic loss rates are used to arrive at the percentage of land required for total disposal.


PUMP
1
2
3
4
5
WELL
TO IRRIGATION
FENCE
ACCESS
BUFFER POND

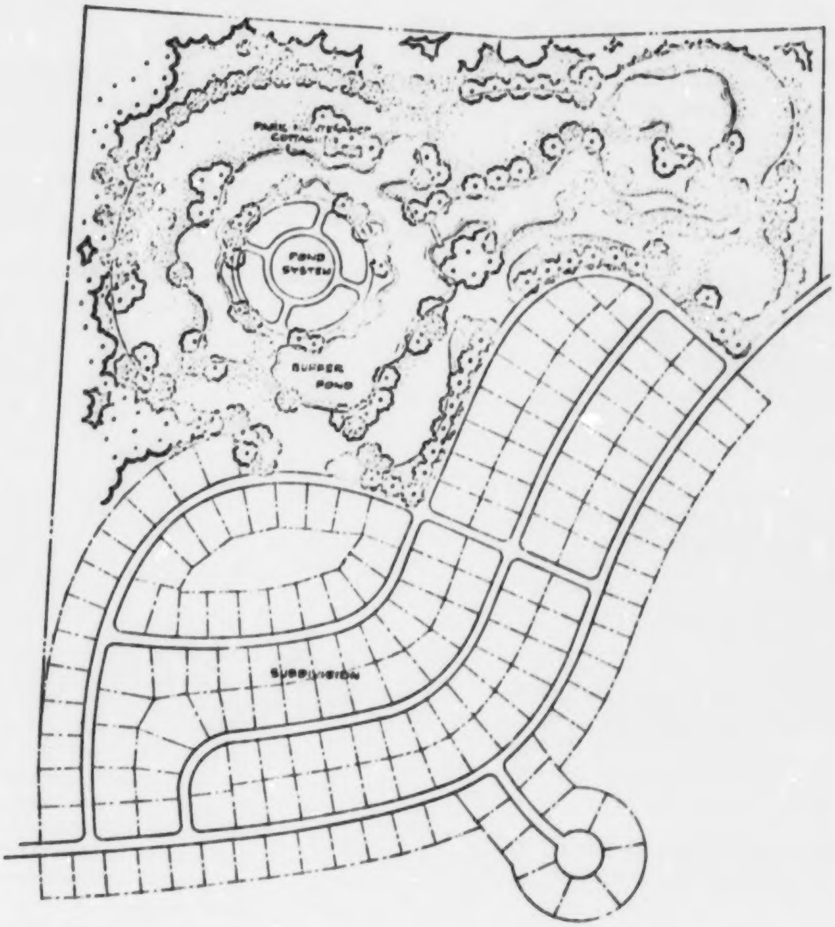
PLAN


PUMP
FENCE
SCUM BAFFLES
WELL PUMP
TO IRRIGATION
WELL

SECTION

FIG. 4 SCHEMATIC - INTEGRATED SERIES PONDS

2 **A typical pond system is outlined in Figure 4 and the position within a subdivision as indicated in Figure 5.**


PARK MAINTENANCE COTTAGE
POND SYSTEM
BUFFER POND
SUBDIVISION

POND SYSTEM IN PARK OF SUBDIVISION DESIGN

56<

3 As pointed out by Noren (3), the park-pond integrated system has several advantages associated with advanced planning. Community sewers are installed in advance in smaller lots than if septic tanks were required. Unit savings in costs of sewers, streets and utilities are significant. Land saved in individual lots may be consolidated into open space held as common use property. Operation and maintenence of the common land uses similar equipment and manpower as park property.

4 The pond complex must be designed for the specific climate soil and population. The integrated pond is more costly than simple ponds. The system is less costly than mechanical plants, however. They require little operational skill but must not be neglected.

D Overflowing Ponds

1 Terminal ponds may be impossible or excessively costly where rainfall greatly exceeds evaporation, percolation is limited, or land is already developed. Intermittent or overflowing ponds may be required.

2 Mass diagrams of predicted flow, evaporation and percolation for seasonal changes must be included to reveal the crucial periods for detention time estimates.

a The critical detention time in days (θ) may be estimated:

$$\theta = \frac{Ad}{Q\text{ in} - A(P_n + E_n)} \quad (4)$$

where A- is expressed in Acres
d-is depth in ft
Q_{in}- in Acre ft per day
$A(P_n + E_n)$-represents the volume loss function of net percolation and evaporation expressed in ft/day.

b Formula 4 shows that detention increases more rapidly than area and very rapidly as the loss fraction approaches Q in.

c Table III includes relationships of design flow, pond, area, depth, and volume, plotted graphically in Fig. 6.

TABLE III
VARIATION OF DETENTION PERIOD WITH INCREASING POND AREA FOR OVERFLOWING PONDS

Design Flow MGD	Assumed Pond Characteristics			Assumed Q_{in} Ac-ft Day^{-1}	P+E Ac-ft Day^{-1}	Q_{out} O'Flow Ac-ft Day^{-1}	θ Det Per
	Area Acres	Depth ft	Vol Ac-ft				
0.77	5	6	30	2.36	0.3	2.06	14.9
0.77	10	6	60	2.36	0.6	1.76	34.1
0.77	15	6	90	2.36	0.9	1.46	61.5
0.77	20	6	120	2.36	1.2	1.16	103.0
0.77	25	6	150	2.36	1.5	0.86	174.0
0.77	30	6	180	2.36	1.8	0.58	310.0
0.77	35	6	210	2.36	2.1	0.26	816.0
0.77	40	6	240	2.36	2.4	0.00	

d The detention is a function of the integrated factors listed and the events taking place. Evaporation and percolation rates differ in ponds that do or do not contain significant algae but experimental data are insufficient to generalize with respect to detention time. Detention time estimates are only as good as the estimates for P_n, E_n, and flow.

E Treatment of Wastes

1 Significant removal of settleable solids, nutrients, oxygen demanding materials, toxicity, hardness, and pathogens occurs in ponds whether this action is intended or not. Each

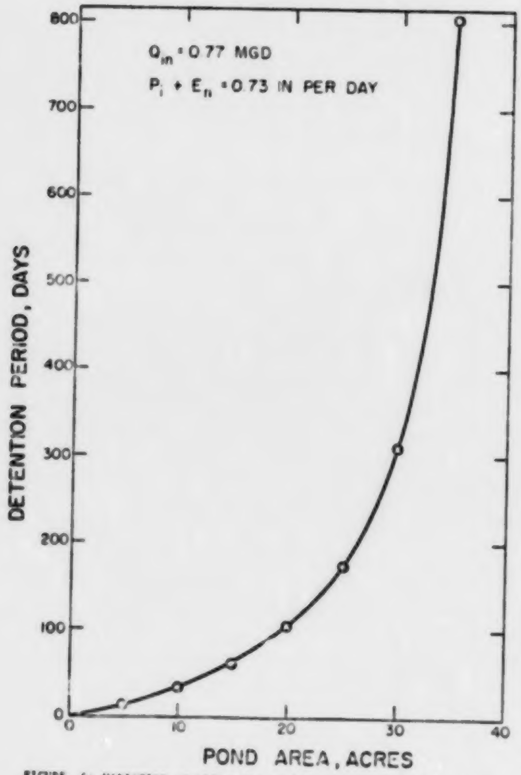


FIGURE 6: VARIATION OF DETENTION PERIOD WITH INCREASING POND AREA FOR OVERFLOWING PONDS

action responds at certain rates that vary with the conditions and reactants. Temperature is a major variable in turn controlled by solar radiation.

a It may be expected that 90% of the settleable material will be removed in 3 days, 99% in 10 days. Growth of algae tends to obliterate the settling record with newly formed suspended solids.

b BOD removal (Table IV) is strongly influenced by temperature in a direct manner. Even at 5°C 90% BOD removal is likely at 100 days. The BOD data may or may not reflect algal activity and newly formed particulates in short time tests.

c Coliform removal is a function of time, temperature and conditions as plotted in Fig. 7. Design criteria on coliform removal at low water temperatures may represent + 200 days detention. Coliform group removal is primarily related to pH(7) and generally is the limiting value compared with removal of other indexes.

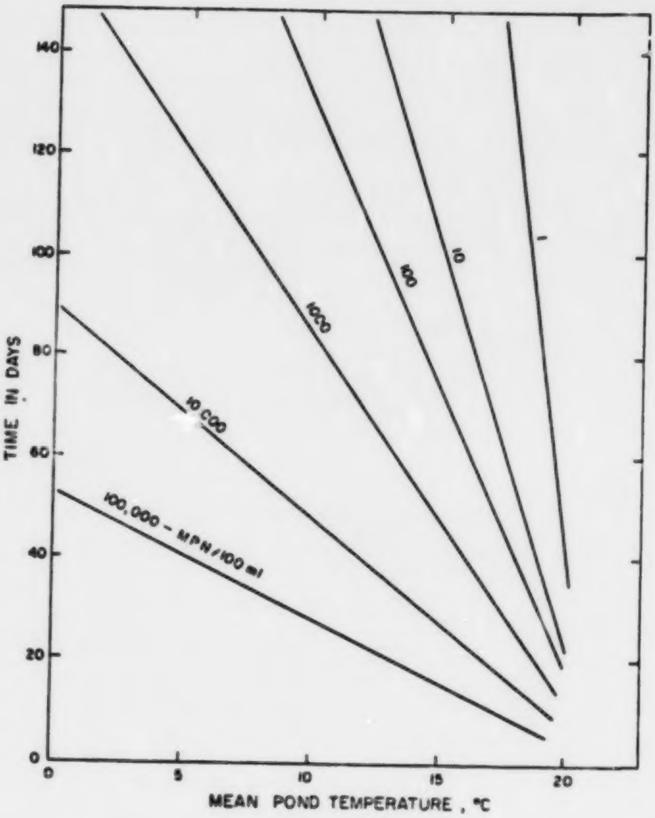


FIGURE 7 OBSERVED MPN/100 ML IN A SINGLE FACULTATIVE POND AS A FUNCTION OF TIME AND TEMPERATURE INITIAL MPN 10^8/ML

TABLE IV

Natural Reduction in Supernatant BOD in an Undisturbed Pond

Time Days	Percent BOD Removal at Indicated Ave. Temp.			
	10°C	14°C	15°C	18°C
0	0	0	0	0
1	23	35	47	33
2	40	61	56	53
4	46	70	67	56
9	57	75	75	81
16	72	89	90	87
32	90	90	95	96
64	94	--	94	--

2 Insolubles formed in a pond as a result of biological oxidation and photosynthesis are mutually interrelated in line with oxygen (algal) and CO_2 (bacterial) production.

a Algae production may be approximated (without serious nutrient deficiences by the following:

$$W_a = 0.15\ F\ S\ T_c \qquad (5)$$

where W_A = lbs of algae/A/day

F = photosynthetic efficiency

S = solar energy/day

T_c = temperature coefficient

b The amount of oxygen produced is about 1.5 times the weight of algae and oxygen needed is about 1.6 times BOD requirements. Oxygen production also can be estimated from Eq5.

c Table V presents solar energy tabulation by month and latitude. Whenever solar energy is below 100 cal/sq cm/day, low or freezing water temperatures are likely and photosynthesis is negligible.

d Non overflow ponds are stressed because algal production releases oxygen during rapid growth. O_2 demand of algal cell mass decay comes later.

e Table VI presents relationships of K_T with temperature for a specific organism strain. General behavior is fairly well established, but specifics are not. This table shows maximum activity but it is not certain that this holds for all situations and organisms.

f Photosynthetic efficiency is highly variable and important in prediction of oxygen production. Mean efficiency in outdoor ponds at 20°C tends to be about 4%. The efficiency of completely mixed cultures declines with increasing light intensity and duration and increases with depth up to a depth of about 3 times the depth of light penetration. Net photosynthesis within a given algae cell is a resultant of two variable functions of photosynthesis and respiration, i.e.,

$$P\ net = (P - R) \qquad (6)$$

TABLE V

PROBABLE VALUES OF VISIBLE SOLAR ENERGY AS A FUNCTION OF LATITUDE AND MONTH

Latitude	Month											
	Jan	Feb	Mar	Apr	May	Jun	Jul	Aug	Sep	Oct	Nov	Dec
0 max	255[a]	266	271	266	249	236	238	252	269	265	256	253
min	210	219	206	188	182	103	137	167	207	203	202	195
10 mas	223	244	264	271	270	262	265	266	266	248	228	225
min	179	184	193	183	192	129	158	176	196	181	176	162
20 max	182	213	246	271	284	284	282	272	252	224	190	182
min	134	140	168	170	194	148	172	177	176	150	138	120
30 max	136	176	218	261	290	296	289	271	231	192	148	126
min	76	96	134	151	184	163	178	166	147	113	90	70
40 max	80	130	181	181	286	298	288	258	203	152	95	66
min	30	53	95	125	162	173	172	147	112	72	42	24
50 max	28	70	141	210	271	297	280	236	166	100	40	26
min	10	19	58	97	144	176	155	125	73	40	15	7
60 max	7	32	107	176	249	294	268	205	126	43	10	5
min	2	4	33	79	132	174	144	100	38	26	3	1

[a] Values of S in cal per sq cm per day.

To determine average value of S: $S_{ave} = S_{min} + p\,(S_{max} - S_{min})$ in which

p = total hr sunshine divided by total possible hrs sunshine.

To determine yield of algal cell material in lb of algal cell material per acre per day:
Y = 0.15 FS. This formula applies to a mean temp. of 20°C.
For other temperatures apply temperature coefficients from Table 5.

To determine yield of oxygen in lb of oxygen per acre per day: $O_2 = 0.25$ FS.

This formula applies to a mean temperature of 20°C. For other temperatures, apply temperature coefficients from Table 5.

60<

TABLE VI

TEMPERATURE COEFFICIENTS FOR CHLORELLA PYRENOIDOSA* GROWN IN SEWAGE

Temp°C	k_T	Temp°C	k_T
0	0.00	18	0.96
2	0.00	20	1.00
4	0.00	22	0.99
6	0.02	24	0.96
8	0.23	26	0.92
10	0.49	2	0.87
12	0.70	30	0.80
14	0.82	32	0.73
16	0.91	34	0.65

*Emersion strain

The cells are the products of (P - R). Few have studied respiration rates of photosynthesizing algae. McCarthy & Forwe (14) found that cells in a dark aerobic environment lost about 3% of dry wt/day. The author observed rates 10 to 20 times that at 25 to 30°C.

3 Nutrient removal

a The conversion of soluble nutrients to removable particulate form is an obvious means of nutrient removal. Off gas CO_2 removal is significant if algal fixation is limited. Methane fermentation, nitrogen oxidation and denitrification, phosphorus and calcium carbonate precipitation are significant. Other mechanisms are possible.

b Detailed discussion of methane fermentation is not intended herein (see 4, 5). The general formula for fermentation gas production in ponds is

$$G = 450\,(T - 15) \qquad (7)$$

where G = gas emission cu.ft./A./day
T = temperature in °C

Gas production occurs below 15°C but temperature dependence is irregular. Emitted gas composition indicates takeout of about 1 lb BOD/10 cu.ft. of evolved gas. Gas emission results in carbon takeout in the form of methane or carbon dioxide.

c Deep facultative ponds are receiving renewed interest because of the takeout of up to 80% of the combined nitrogen content in process. This is the cheapest available method for nitrogen removal. Solids removal accounts for part of this. Nitrification of unoxidized nitrogen in aerobic zones of the pond followed by denitrification in deeper waters of the pond results in release of molecular nitrogen. The nitrate oxygen is used for respiratory demand in DO deficient water. Other mechanisms may be included but the two mentioned are significant.

d Phosphorous removal by precipitation as a hydroxy phosphate occurs in high pH waters. Mineral content of the incoming water influences phosphate conversion and form. Biological factors including luxury uptake of P on cell mass are important. It is not possible to generalize precise behavior of P in a complex and changing situation. Recycle characteristics of P complicate interpretation and prediction.

e Sobsey (8) observed that the presence of algae favored an increased removal of viruses in stabilization ponds.

f Klein and Kaufman (9) reported almost complete destruction of NTA.

g Reclamation aspects of ponds with respect to nutrient water and land are well documented by Dryden and Stern (10), Oswald (11 & 12) and McGarry (13).

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CHEMICAL-PHYSICAL PROCESSES

I INTRODUCTION

A The most recent advance in the art of sewage treatment is the application of physical-chemical technology to raw sewage treatment. In reality, physical-chemical treatment of sewage is not a completely new technology. Chemical treatment of sewage was widely practiced in England and the United States in the latter portion of the 19th Century. However, this technique gradually fell into disuse with the advent of activated sludge, because activated sludge systems achieved higher degrees of treatment (1). In the 1930's, a number of systems employing physical and chemical treatment in combination were evaluated. These produced treatment superior to primary sedimentation followed by activated sludge (conventional treatment) but at a cost 1.5 to 2 times as great (2). This additional increment of cost discouraged adoption of physical-chemical treatment at that time.

B During the last decade, advances in physical-chemical treatment technology resulting from Environmental Protection Agency-supported research have significantly reduced the cost of physical-chemical treatment. In addition, it is apparent that higher levels of treatment will be required in the future to maintain water quality. As a result of these alterations in conditions, physical-chemical treatment is now an alternative to conventional treatment, especially for situations where significant phosphorus removal is required.

II GOALS OF SEWAGE TREATMENT

A The fundamental goal of sewage treatment is sufficient reduction in the level of pollutants in the wastewater to allow discharge to the environment. The pollutants in sewage are grouped into classes of similar compounds which have the same environmental impact rather than being individually dealt with. At present, the five major pollutant groups of interest are: 1) suspended solids, 2) organic matter as measured by BOD, TOC or COD, 3) phosphorus compounds, 4) nitrogen compounds, and 5) pathogenic microorganisms.

B It is impossible to specify one set of effluent standards which is applicable to all or most situations. However, it is generally agreed that a good quality effluent will have the characteristics given in Table 1. In addition, complete disinfection will be required. As yet, no general agreement on a nitrogen level is in evidence. It must be repeated the goals listed above are not design standards, but are guides in describing a good quality effluent.

TABLE 1.

DESIRED EFFLUENT QUALITY	
BOD	10 MG/L
COD	60 MG/L
SS	10 MG/L
P	1 MG/L

III THE CLARIFICATION-ADSORPTION PROCESS

A The physical-chemical treatment system which at present is most advanced in development, and which seems the best is the clarification-adsorption system schematically illustrated in Figure 1.

Figure 1.

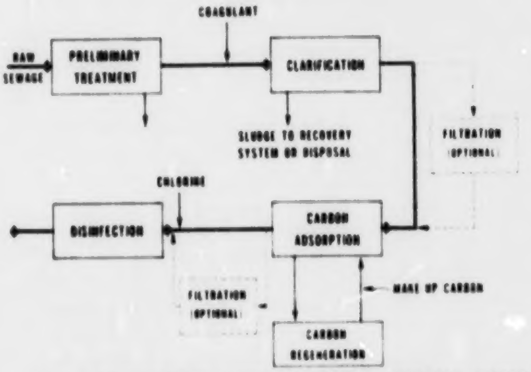


FIG.1 FLOW DIAGRAM OF A PHYSICAL CHEMICAL TREATMENT SYSTEM

In this system, chemical clarification and filtration are utilized to achieve almost complete suspended and colloidal solids removal. With a high enough dose of the proper chemicals, essentially complete phosphorus removal can also be obtained. The function of the activated carbon step is the removal of soluble organics.

As in any conventional system preliminary screening, grit removal and final disinfection are provided. The filtration step is shown as optional but conservative design dictates its use. The positioning of this step prior to or after carbon adsorption is dictated by the type of waste-carbon contacting system provided. The three steps of chemical clarification, filtration and carbon adsorption will be discussed in some detail below.

This system does not provide ammonia or nitrate-nitrogen removal. Physical-chemical methods for soluble nitrogen removal are covered in another lecture of this symposium. Any of these nitrogen removal methods can be integrated into the clarification-adsorption system.

B Chemical Clarification

1 Chemical clarification provides the bulk of the pollutant removal achieved by this type of treatment system. In this unit operation an appropriate chemical is dosed to the wastewater and the mixture is flocculated and settled. All of these steps can be conducted in independent units, or combined into a single unit which is usually referred to as solids-contact clarifier.

2 Chemicals which have been successfully used to clarify raw sewage include organic polymers (3), iron salts (4), aluminum salts (5), and lime (6) (7). The inorganic coagulants have the advantage of providing for phosphorus removal. At present there is no rational method for predicting the dose of chemical required. For planning purposes, jar tests are suggested. Fortunately, field control of the coagulant dose is possible. For all coagulants except lime the suggested method is monitoring of the clarifier effluent turbidity. Good clarification is always obtained when sufficient chemical is added to provide good phosphorus removal. With lime as the coagulant, excellent control is obtained by pH measurement. The pH required to achieve good clarification and phosphorus removal is dependent on the chemical characteristics of the sewage. In areas where the alkalinity and hardness are low, the high lime process in which the pH is raised to approximately 11.5 is required. Generally, two-stage precipitation with intermediate recarbonation is utilized. In hard water, high alkalinity areas a low lime single-stage precipitation at pH 9.5 to 10 is sufficient.

Table 2 illustrates the treatment obtained with chemical clarification in pilot plants at a variety of locations. In addition to the expected high levels of suspended solids and phosphorus removal significant organic removal was obtained. Based on these data and others it can be expected that chemical clarification of raw sewage will consistently yield organic removal in the 65-75 percent range. With this degree of organic removal from chemical coagulation the carbon need provide only a small increment of additional removal to match the performance of a good secondary treatment system.

TABLE 2
ACHIEVEMENTS OF CHEMICAL CLARIFICATION

PLANT	CHEMICAL	ORGANIC REMOVAL %	SS REMOVAL %	P REMOVAL %
EWING-LAWRENCE	170 mg/l $FeCl_3$	80	95	90
NEW ROCHELLE (ZM)	LIME pH 11.5	80	98	98
WESTGATE, VA.	125 mg/l $FeCl_3$	70	-	-
SALT LAKE CITY	80-100mg/l $FeCl_3$	75		80
BLUE PLAINS	LIME pH 11.5	80	90	95

3 Sludge disposal plays a dominant role in the economics of any chemical clarification system. Only limited data on the characteristics of sludges resulting from the chemical treatment of raw sewage are available. These data indicate that with iron or aluminum salts as coagulants:

a The resulting sludge volume sometimes exceeds and sometimes is less than primary sludge from the same sewage.

b The chemical sludge is more difficult to dewater in a vacuum filter than the corresponding primary sludge.

c With lime as the coagulation chemical dewatering is extremely rapid just by gravity thickening. Eimco (8) reported lime-raw sewage sludge solids concentrations of 15% to 25% after gravity thickening.

4 In addition to yielding a sludge amenable to rapid and easy dewatering, lime is the only coagulation chemical which can be economically recovered. Recalcining the lime sludge in a furnace has been successfully conducted in a tertiary plant at South Lake Tahoe and has been state-of-the-art in water softening plants for a considerable period of time (9). In addition to regenerating the lime, the organic solids are incinerated, thereby accomplishing ultimate disposal of the sludge. For these reasons, it is anticipated that lime will be the coagulant of choice for most situations.

5 Design criteria for the coagulation equipment is similar to that used in water treatment plants. A flash mix of one minute, flocculation for 15 to 30 minutes and sedimentation at upflow rates of 0.5 and 1.0 gpm/ft^2.

C Carbon Adsorption Systems

1 The role of the carbon adsorption step is the removal of soluble organics from the wastewater after coagulation has removed particulates. The total organic removal achieved by the combination of clarification and carbon adsorption at several physical-chemical pilot plants is illustrated in Table 3. It can be seen that not only is the removal achieved quite high (95%+) but equally important the residual organics after treatment are quite low. These effluents are superior to that listed in Table 1.

Table 3.

TOTAL TREATMENT P-C PLANTS

PLANT	ORGANIC REMOVAL %	EFFLUENT CONCENTRATION
BLUE PLAINS	95-98	TOC = 6
LEBANON	95	TOC = 11
EWING-LAWRENCE	95-98	TOC = 3-5
NEW ROCHELLE	95	COD = 8

2 The results reported in Table 3 were obtained with carbon contacting systems employing granular carbon. In such systems, the waste is passed either upward or downward through columns containing the carbon. Downflow columns function as packed beds and accomplish filtration of the wastewater. Flow rates of 2 gpm/ft^2 to 8 gpm/ft^2 have been employed. In this flow range essentially equivalent adsorption efficiency is obtained provided the same contact time is employed. At flow rates below 2 gpm/ft^2 adsorption efficiency is reduced, while at flow rates above 8 gpm/ft^2 excessive pressure drop takes place. Contact times employed are in the range of 30 minutes to 60 minutes on an empty bed basis. In general increases in contact time up to about 30 minutes yield proportionate increases in organic removal. Beyond 30 minutes the rate of increase falls off with increases in contact time becoming negligible at 60 minutes. Gravity beds are commonly used for lower flow rates, pressure vessels for higher flow rates. The pressure vessel is more expensive but requires less land and accepts flow variations better.

3 Provision must be made to periodically backwash downflow carbon beds because they gradually collect suspended solids. In addition, biological growth takes place on the carbon granules and tends to clog the bed. It is advisable to include a surface wash and air scour to be assured of removal of the gelatinous biological growth.

4 Backwash of the carbon beds satisfactorily relieves clogging but does not completely remove the biological growth. Significant biological activity is manifest in the carbon beds at most times. This leads to the development of anaerobic conditions in the carbon bed and generation of sulfides. Aeration of the column feed has been utilized to prevent anaerobic conditions, but this produced so much biological growth that excessive backwash was required.

5 In an attempt to overcome these difficulties, upflow carbon columns have been operated in a slightly expanded mode (about 10% expansion). This allows for significant accumulation of biological activity on the carbon granule with little increase in head loss. Consequently, aerobic conditions can be maintained and sulfide generation prevented.

6 Backwash facilities are necessary for expanded beds to remove excess growth. Flow rate range which can be used is more restricted than with packed bed systems. With the commercial sizes of carbon available, (8x30 mesh or 16x40 mesh) flow rates above 4 to 5 gpm/ft^2 are required to achieve the proper degree of expansion. Care must be exercised to avoid hydraulic surges which could wash carbon out of the system. The filtration section of the physical-chemical treatment plant must follow carbon contact when the expanded bed system is used.

7 The latest carbon contacting scheme to be applied to waste treatment utilizes powdered carbon (particle size below 200 mesh). This procedure provides for a mixture of a carbon slurry and the wastewater in a Reactor-Clarifier. Polymer addition is generally required to achieve a good gravity separation of the carbon from the wastewater following contact. The potential advantages of this sytem are the use of a cheaper carbon (10¢ per pound vs. 30¢ per pound for granular carbon) and a simpler type of contacting system.

8 A critical aspect of the design of any carbon contacting system is the expected capacity of the carbon for the organics. In the chemical processing industries, this is evaluated by running

adsorption isotherm tests. In the waste treatment field, isotherms are of limited utility because the biological activity which develops on the carbon tends to greatly enhance its apparent capacity for organic removal.

D Design of Carbon Adsorption Facilities

1 The role of the design engineer is to utilize the greatest use of the carbon regardless of capacity. Counter-current contact favors good effluents and carbon utilization. This is achieved by having the waste flow through a number of contactors or stages in series in one direction, while the carbon moves in the opposite direction. In the powdered carbon contacting system, this is exactly the procedure used. With granular carbon, this procedure cannot be used as undesirable attrition losses will take place. When an undesirable effluent is obtained, the lead contactor is removed from service and a spare contactor with fresh carbon placed at the end of the line. Each contactor is then moved up one position in the line. This is accomplished by piping and valving a series of columns to shift the inflow and outflow points rather than physically moving the columns. As the number of stages increases, the piping and valving arrangement becomes more complex and costly. In design, a compromise between the advantage of adding another stage to more closely approach the highest use of the carbon capacity, and the cost of each additional stage must be achieved.

2 An alternate arrangement provides parallel flow through a number of identical contactors. Each is at a different stage of exhaustion and produces a slightly different effluent quality. These individual effluents are blended to produce the final product.

Table 4 reviews the various factors which must be considered in designing any carbon removal system.

Table 4.

CONSIDERATIONS IN CARBON TREATMENT DESIGN

1. TYPE OF CARBON-GRANULAR OR POWDERED
2. CONTACT TIME
3. FLOW RATE
4. CONFIGURATION-SERIES OR PARALLEL
5. NUMBER OF STAGES
6. FLOW DIRECTION - PACKED OR EXPANDED
7. HYDRAULIC FORCE - PUMPED OR GRAVITY
8. ORGANIC CAPACITY

3 Carbon capacities obtained in field operation at various pilot plants are given in Table 5. The waste, effluent criteria, number of contact stages, etc., varied from plant to plant giving some spread in the results. These capacities are expressed as pounts of organics removed (either as COD or TOC) per pound of carbon. For general planning purposes, a capacity of 0.5 pounds of COD per pound of carbon is reasonable. This is approximately equivalent to a requirement of 500 pounds of activated carbon per million gallons of sewage treated.

Table 5.

Carbon Capcity in P-C Treatment Plants

Plant	Carbon Capacity	
	LBS TOC / LB A.C.	LBS COD / LB A.C.
Blue Plains	0.15	0.41
Ewing-Lawrence	0.3	-
New Rochelle (ZM)	-	0.6
Lebanon	0.22	0.5

AC = Activated Carbon

4 A carbon requirement at this level would be prohibitively expensive if regeneration and reuse of the exhausted carbon were not possible. At present, a technically and economically feasible method is available for the regeneration of granular activated carbon. This method requires heating of the carbon in a multi-hearth furnace in the presence of steam to 1750°F. This treatment burns away the adsorbed and trapped organics. During a regeneration cycle, some of the carbon is physically lost by burning and attrition, and some of each particle's capacity is lost by alteration of surface properties. The overall loss expressed as percent by weight of virgin carbon required to restore the total original capacity of the batch ranges from 5 to 10 percent (10). Thus, for planning purposes carbon makeup requirements at a plant should range from 25 to 50 lbs per million gallons of sewage treated.

5 At present, experiments on a fluidized bed regeneration system for powdered carbon are moving into the large-scale pilot stage. Successful completion of these tests will be a big step forward in making a powdered carbon system a technical and economic reality. The key factor will be maintaining the carbon loss at a low enough level during the regeneration.

IV DESCRIPTION AND PERFORMANCE OF SPECIFIC PHYSICAL-CHEMICAL PILOT PLANTS

A Ewing-Lawrence (4)

For the last several years, FMC Corporation has been conducting studies of physical-chemical treatment at the Ewing-Lawrence Treatment Plant near Princeton, N.J. Figure 2 illustrates the clarification section of the plant. These units were somewhat overdesigned to insure a good feed to the adsorption section of the plant, as the major purpose of the research was to evaluate carbon performance. Figure 3 is a schematic of the overall plant. Packed bed versus expanded bed carbon contact in parallel was studied at this plant. Figure 4 indicates that the performance of the packed bed contactor was slightly superior to that of the expanded bed contactors. However, the investigators indicated that the advantage was not sufficient to offset the need to backwash more frequently and the potential difficulties from anaerobic conditions in the carbon beds. Figure 5 illustrates the BOD removal obtained over a 3-1/2 month period. It can be seen that despite considerable variation in influent strength a consistently good effluent was produced. This figure also reiterates that the bulk of the treatment is accomplished in the chemical clarification step.

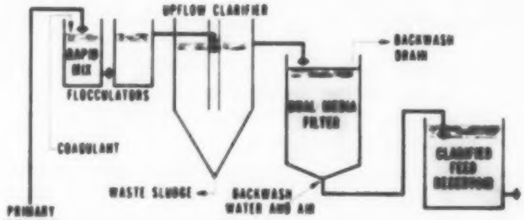
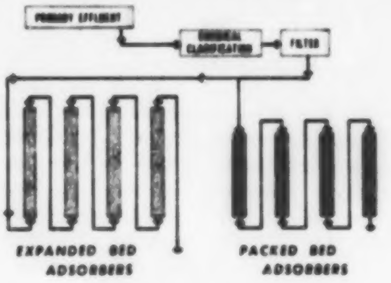


FIG 2

FLOW DIAGRAM OF CLARIFICATION SYSTEM

FIG. 3
EXPERIMENTAL SET-UP
FOR 24 FT. CARBON BEDS



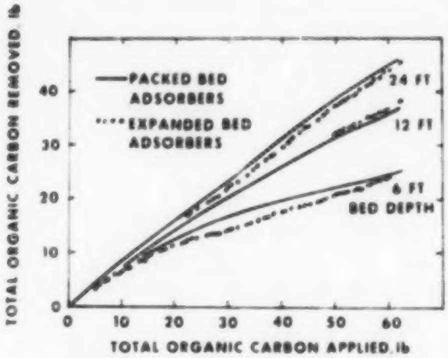


FIGURE 4 Cumulative sorption of total organic carbon for packed bed and expanded - bed adsorbers as a function of TOC applied.(Lb × 0.454 = kg. ft × 0.3 = m.)

B Blue Plains (7)

At the Blue Plains treatment plant in Washington, D. C., a pilot physical-chemical treatment system is being operated as part of a joint effort of EPA and the District of Columbia. Figure 6 is a schematic of the treatment system. Two-stage high lime clarification is provided with a small dose of ferric chloride in the second stage for flocculation of calcium carbonate. This stage of the plant has a capacity of 100, 000 gal/day. The filtration, ion exchange and downflow two-stage carbon beds have a capacity of 50, 000 gal/day. This plant is highly automated and thus runs with a minimum of manual control. For example, the lime dose is automatically set by monitoring the pH and alkalinity in the clarifiers, and backwash of the filters, ion exchange columns, and carbon columns is automatically initiated by pressure drop and/or a time cycle.

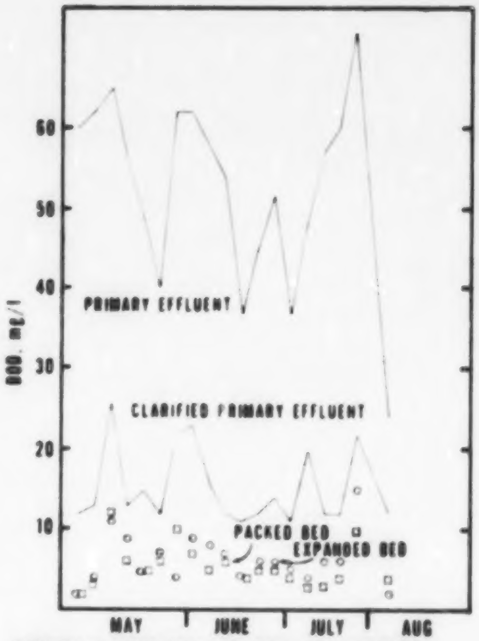


FIGURE 5 REMOVAL OF BOD BY CHEMICAL CLARIFICATION AND ACTIVATED CARBON

Figure 7 illustrates the treatment performance obtained at this plant for a variety of pollution control parameters. Nitrogen removal was obtained by the use of an ion exchange process with a zeolite which has a high affinity for ammonium ion. Again, these data indicate that the bulk of the work is done by the clarification step with the filter and carbon columns serving as polishing devices. Table 6 gives the average effluent characteristics for this installation. This effluent is far better in quality than normal secondary effluent.

FIG. 6
INDEPENDENT PHYSICAL CHEMICAL TREATMENT
LIME PRECIPATION

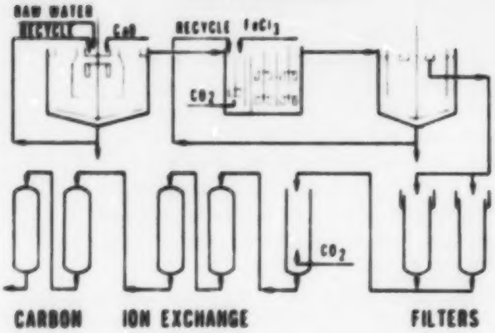
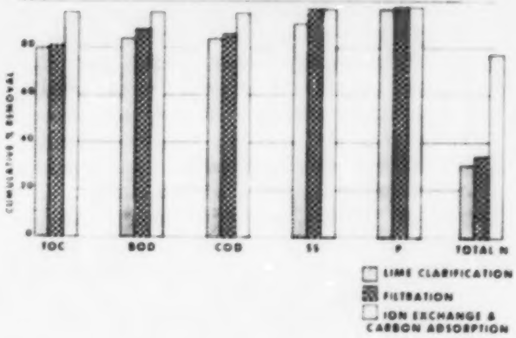


TABLE 6

EFFLUENT QUALITY - P - C
PILOT PLANT AT BLUE PLAINS

TOC	6 mg/l
BOD	5 mg/l
COD	13 mg/l
SS	5 mg/l
TOTAL P	0.15 mg/l
TOTAL N	4.6 mg/l

FIG. 7 REMOVAL EFFICIENCIES
PHYSICAL CHEMICAL TREATMENT
BLUE PLAINS PILOT PLANT



C Salt Lake City, Utah (8)

Under contract between EPA and Eimco Corporation, an evaluation is being conducted of a physical-chemical treatment system employing powdered carbon contact after chemical clarification. Figure 8 illustrates the pilot plant. Two countercurrent stages of carbon contact in Reactor-Clarifier are provided following chemical clarification. Some results with lime as the coagulant are given in Table 7. These results are not quite as good as in the plants using granular carbon discussed previously. The effluent, however, must still be classified as of good quality. One potential advantage of this contacting system is the ability to pace the carbon dose to the organic demand. Evaluation of powdered carbon regeneration will take place at this installation. The spent carbon will be dewatered on a vacuum filter prior to being fed to the fluid bed furnace.

TABLE 7

Powdered Carbon Pilot Plant
Operating Conditions

Flow Rate 50 Gal/Min
Chemical 425 mg/l Lime TO pH 10.8
Carbon 150 mg/l + 0.4 mg/l polymer

RESULTS

	COD	BOD	SS	P
Raw Sewage	222	144	200	7.3
Clarified Effluent	65	47	28	1.4
Final Effluent	35	13	7	0.4

ALL RESULTS GIVEN IN mg/l

D **Rocky River, Ohio** (3)

Several studies have been made utilizing polymer addition to existing primary plants followed by small-scale pilot carbon adsorption. One such study was done at the 10 MGD primary facility at Rocky River, Ohio. An anionic polymer was added to the existing primary clarifier at a dosage of 0.3 mg/1, and a side stream of clarified effluent was applied to small carbon columns for a period of about one month. The summary of the data is shown in Table 8. It can be seen that, even with less than optimum clarification, effluent comparable to good secondary effluent was produced with 33 minutes carbon contact time. On the basis of this preliminary work, the City of Rocky River applied for and was awarded a Research and Development Grant from EPA to help support a full-scale investigation of the clarification-adsorption process for secondary treatment.

1 One of the principal motivations for the city to use physical-chemical treatment is shown in Figure 9. The installation of conventional activated sludge facilities would necessitate the condemnation of a considerable area of expensive property, whereas a carbon adsorption system could easily fit into the existing site. Testing has been undertaken to determine what coagulant or combination of chemicals will be used in the clarification system. If phosphorus removal is required, an inorganic coagulant will necessarily be the choice.

FIGURE 8

PHYSICAL/CHEMICAL TREATMENT PROCESS

EIMCO
ENVIROTECH

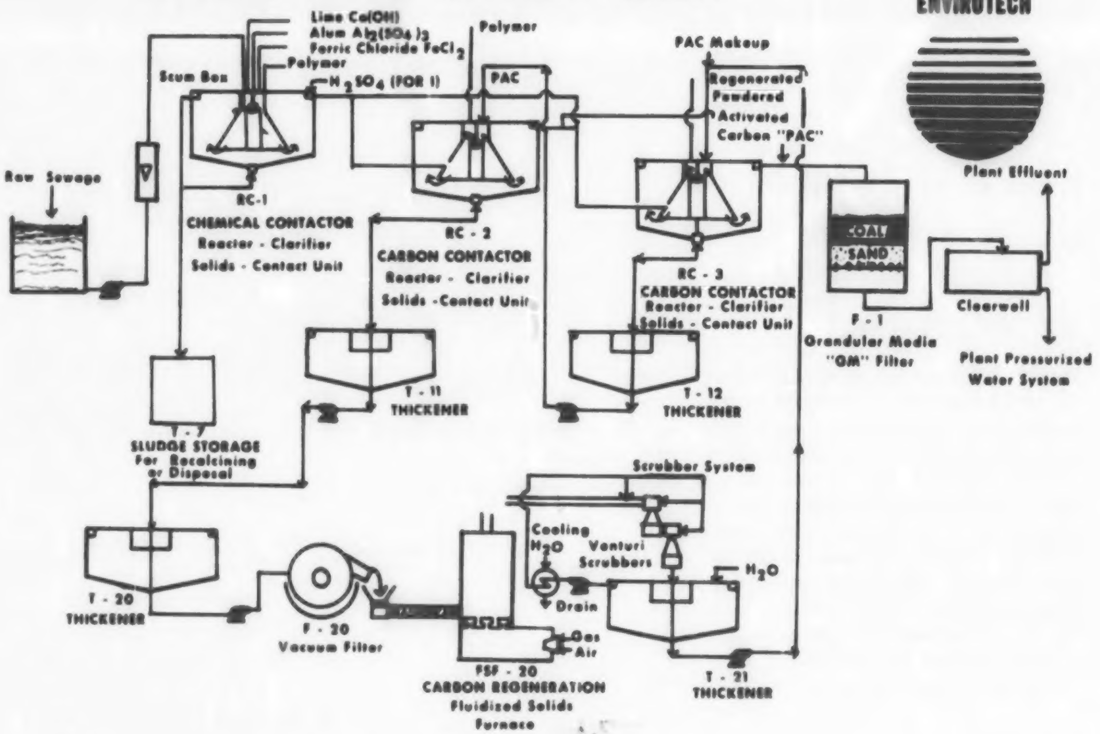

Lime Ca(OH)
Alum $Al_2(SO_4)_3$
Ferric Chloride $FeCl_2$
Polymer
Polymer
Scum Box
H_2SO_4 (FOR I)
PAC
PAC Makeup
Regenerated Powdered Activated Carbon "PAC"
Raw Sewage
RC-1
CHEMICAL CONTACTOR
Reactor - Clarifier
Solids - Contact Unit
RC - 2
CARBON CONTACTOR
Reactor - Clarifier
Solids - Contact Unit
RC - 3
CARBON CONTACTOR
Reactor - Clarifier
Solids - Contact Unit
COAL
SAND
F - 1
Granular Media "GM" Filter
Clearwell
Plant Effluent
Plant Pressurized Water System
T - 11
THICKENER
T - 12
THICKENER
T - 7
SLUDGE STORAGE
For Recalcining or Disposal
Scrubber System
Cooling H_2O
Venturi Scrubbers
H_2O
Drain
T - 20
THICKENER
F - 20
Vacuum Filter
Gas
Air
FSF - 20
CARBON REGENERATION
Fluidized Solids Furnace
T - 21
THICKENER

Figure 9.

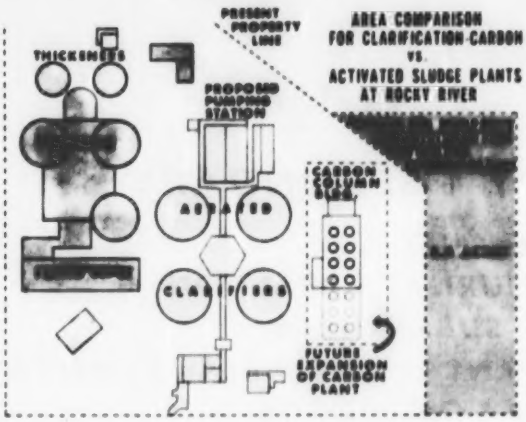

PRESENT PROPERTY LINE
AREA COMPARISON FOR CLARIFICATION-CARBON VS. ACTIVATED SLUDGE PLANTS AT ROCKY RIVER
THICKENERS
PROPOSED PUMPING STATION
CARBON COLUMN BLDG.
AERATED
CLARIFIERS
FUTURE EXPANSION OF CARBON PLANT

TABLE 8

ROCKY RIVER
WASTE TREATMENT PLANT
CLARIFICATION-CARBON PROCESS

	Raw	Polymer Clarification	Carbon Contact Time, Minutes			Percent Removed
			14	23.4	32.6	
SUSPENDED SOLIDS, mg/l	107	65	13	15	7	93.3
BOD, mg/l	118	57	21	11	8	93.3
COD, mg/l	235	177	67	50	44	81.3

V ADVANTAGES OF PHYSICAL-CHEMICAL VS. CONVENTIONAL TREATMENT

Perhaps the most important advantage is the stability of operation provided by a treatment system based on physical and chemical technology. Biological systems are notoriously sensitive to changes in environmental conditions. If a toxic material or a hydraulic peak occurs, not only will the efficacy of the biological plant drop off but recovery may take several days to several weeks. In a physical-chemical plant, the filtration system backs up the clarifier and the carbon system backs up the first two; thus, upsets should be unlikely. In addition, it can be expected that an immediate recovery of the plant will take place once the source of upset is eliminated, or corrections are applied. This inherent stability of performance is also reflected in greater design and operational flexibility. Whole sections of a physical-chemical plant can be cut in or out of the process stream as required, and a temporary overload can be absorbed with little effect. A list of many of the major advantages of a physical-chemical system is given in Table 9. Most have been discussed at some point in this paper.

VI COST ESTIMATES

One of the uncertain factors concerning physical-chemical treatment is the cost of the process. Definitive data will not be available until large-scale plants have been built and are in operation for several years. Even then, local conditions may significantly affect the actual costs. Smith (11) has taken the available information from pilot plants and preliminary designs of several proposed large-scale plants and has made cost estimates for various size plants referenced to October 1970. Amortization was 6 percent for 24 years. These are presented in Table 10. Note that ranges are given for each plant size. These ranges represent the spread of data available. As a comparison Smith's estimate for primary and secondary treatment with sludge incineration is 16.5 cents per 1000 gallons at the 10 MGD level. With the addition of single stage lime for phosphorus removal, the cost would rise to 23.5 cents per 1000 gallons which is essentially the same as physical-chemical treatment.

TABLE 9

ADVANTAGES OF PHYSICAL-CHEMICAL TREATMENT VS. CONVENTIONAL PRIMARY + SECONDARY

1. LESS AREA REQUIREMENT-½ TO ¼
2. LOWER SENSITIVITY TO DIURNAL VARIATION
3. NOT AFFECTED BY TOXIC SUBSTANCES
4. POTENTIAL FOR SIGNIFICANT HEAVY METAL REMOVAL
5. SUPERIOR REMOVAL OF P COMPOUNDS
6. GREATER FLEXIBILITY IN DESIGN AND OPERATION
7. SUPERIOR ORGANIC REMOVAL

TABLE 10

PRELIMINARY COST ESTIMATE PHYSICAL - CHEMICAL TREATMENT

Total Amortization + O&M, cents per 1000 gallons

PLANT SIZE, MGD	5	10	100
CHEMICAL CLARIFICATION *	9.5 - 13.5	7.3 - 9.6	4.0 - 5,3
CARBON ADSORPTION	11.5 - 18.0	9.1 - 13.5	4.5 - 7.8
FILTRATION	2.9 - 4.5	2.1 - 3.3	1.0 - 1.4
TOTAL	23.9 - 36.0	18.5 - 26.4	9.5 - 14.5

* Two - stage lime recalcination of sludge

75<

TABLE 11
PHYSICAL CHEMICAL TREATMENT PLANTS

SITE	CAPACITY	STATUS
Rocky River, Ohio	10	Bids taken
Painesville, Ohio	5	Design
Cleveland, Ohio	10	Design
Fitchburg, Massachusetts	15	Design
Waterford, New York	2	Planned
Cortland, New York	10	Planned
Clay, New York	10	Planned
Niagara Falls, New York	60	Planned
Garland, Texas	30	Design
Owosso, Michigan	6	Design

FUTURE DEVELOPMENTS

As a result of the advantages of physical-chemical treatment discussed above and the favorable economic comparison, a number of full-scale treatment plants are being planned or designed. These are listed in Table 11. Only the first two, Rocky River and Painesville, Ohio, will receive Federal research funds. All the others are being planned with no expectation of Federal funds other than the normal construction grant.

This outline was prepared by I. J. Kugelman and J. M. Cohen, Sanitary Engineer and Chief respectively, Physical-Chemical Treatment Research Program, Advanced Waste Treatment Research Laboratory, RATWRC, NERC, EPA, Cincinnati, OH 45268.

SECTION II BASIC PRINCIPLES OF UNIT PROCESSES

	Outline Number
Basic Principles of Coagulation	7
Basic Principles of Carbon Adsorption	8
Basic Principles of Ion Exchange	9
Electrodialysis	10
Reverse Osmosis	11
In-Depth Filtration	12
Microstraining	13
Process Instrumentation	14

BASIC PRINCIPLES OF COAGULATION

I INTRODUCTION

A The natural capacity of our surface waters to assimilate wastes is being increasingly taxed in many areas and even exceeded in some. Methods to treat these wastes to decrease pollution loads must be developed to preserve our natural water resources. A great variety of processes are capable of reducing this pollution, and all of these are currently being carefully examined. An important process is coagulation and flocculation, the subject of this lesson.

B Coagulation is almost exclusively a process for the removal of colloidal and suspended solids. It has application in many areas, such as:

1 Removal of BOD and COD associated with suspended solids from either primary or secondary waste treatment.

2 Removal of phosphate compounds, utilizing both precipitation and coagulation.

3 As a pretreatment process for in-depth filtration.

4 As a clarification step to precede such other processes as electrodialysis, reverse osmosis, carbon adsorption and possibly others.

5 As a concurrent process with flotation or micro-air flotation.

C Each of the above applications imposes different product quality standards which require individual design, mode of operation and economics.

II HISTORY OF COAGULATION

A The principles, methods, and materials for clarifying water have an ancient heritage dating back to the earliest recorded date of about 2000 B.C.

B Methods described by Sanskrit and Egyptian inscriptions demonstrated a knowledge and use of such processes as, sedimentation, coagulation by both organic as well as inorganic compounds, filtration through sand and porous vessels, and disinfection obtained from such metallic containers as copper or bronze.

C The process of coagulation followed by filtration, as we now recognize them, really had their inception in the late 19th Century.

III CURRENT STATUS

A Coagulation and flocculation, generally followed by filtration, is by far the most widely used process for the production of potable water from turbid surface water supplies.

B In the past, domestic wastes were rarely treated by coagulation. Industrial waste treatment frequently required coagulation.

C Use of coagulation and flocculation will predictably be more commonly used in the future: removal of BOD in suspended solids, removal of phosphates, pretreatment for filtration and other processes.

IV DEFINITION

Historically, the terms coagulation and flocculation have been used indiscriminately to describe the process of removal of suspended solids from water. There is, however, a clear distinction between the two terms.

A Coagulation is the destabilization of a dispersion by reducing the forces tending to keep the particles apart. Operationally, coagulation is achieved by adding the appropriate chemical which 1) reduces the forces of repulsion, and 2) increases the number of particles and hence increases probability of particle to particle contact.

B Flocculation is the formation of large or settleable particles from the destabilized colloidal sized particles. In contrast to coagulation, where the primary forces are electrostatic or interionic, flocculation is achieved by particle contact, adsorption, bridging and physical enmeshment mechanisms.

V NATURE OF PARTICLES

A Particles in wastewater may be broadly classified according to their origin as inorganic mineral matter or organic carbonaceous material.

B Particles may additionally be classified according to their size which may range from molecular dimensions to 50 microns or larger. The fraction ≧ 10 microns will settle out given sufficient time. Particles ≦ 1 micron, which are classified as colloidal, will remain suspended for very long times.

C Colloids possess the distinguishing characteristic of high ratio of surface area to mass. For large particles, where this ratio is low, mass effects such as sedimentation under gravity predominate. For colloidal materials, where ratio is high, the properties associated with the surface of the particle, such as electric charges, and ionogenic groups, become more important.

D Colloids may be further classified as hydrophobic and hydrophilic.

E A partial list of particles in wastewater would include:

1 Organics such as polysaccharides, proteins, microorganisms, etc.

2 Inorganic minerals such as clays, silica

3 Inorganic precipitates, hydroxides, metal oxides.

VI STABILITY AND INSTABILITY FACTORS

A Stability refers to the inherent property of colloidal particles to remain dispersed despite passage of time.

B Instability describes the tendency of particles to coalesce whenever particle to particle contact is made.

C For hydrophilic colloid systems, stability is maintained by the phenomenon of hydration in which water molecules are attracted to the surface of particles and act as a barrier to contact between particles.

D Stability of hydrophobic particles is due to the phenomenon of the electrical double layer.

E Each particle suspended in water consists of several components in addition to the primary particle. At least three distinct layers can be described.

F A consequence of the electrical double layer is to create regions of electrical potential in a bulk solution that nominally has a zero potential. The most interesting of the several potentials is the Zeta potential.

G Electrical double layers prevent close approach of particles to each other. Both the thickness of the layer and magnitude of the surface charge and density are important to stability.

VII INSTABILITY FORCES

A Forces which operate in the opposite direction to cause particles to be attracted to each other are called instability forces.

B Brownian motion, inherent to all colloidal particles, promotes instability by causing particle contacts. Brownian motion is important only for particles with diameters 100 millimicrons or less.

C A force which is always one of attraction between particles is Van der Waals force.

This force is operative only within short distances of the order of atomic dimensions, but may be greater for colloidal particles.

VIII MECHANISM OF COAGULATION

A Objective of coagulation is to destabilize the colloids by:

1 Addition or presence of electrolytes which tends to decrease the thickness of the double layer.

2 Addition of ions of opposite charge which tends to neutralize the charge on the particle. The coagulating power of an ion increases sharply with increase of valence.

3 Colloids of opposite charge will cause a coagulation termed mutual coagulation.

4 Coagulation may be obtained by "bridging" such as occurs with polymeric flocculants (polyelectrolytes).

5 Addition of metal coagulants such as alum or iron salts.

IX CHEMISTRY OF METAL COAGULANTS

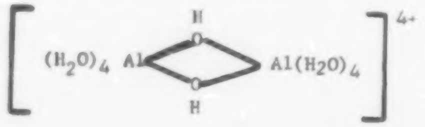
A Coagulation by aluminum and iron salts is not obtained by ionized tripositive aluminum or iron ions but by their hydrolysis products.

B Aluminum salts react with water in a series of chemical reactions described as hydrolysis.

$$[Al(H_2O)_6]^{3+} + H_2O = [Al(H_2O)_5OH]^{2+} + H_3O^+$$

$$[Al(H_2O)_5OH]^{2+} + H_2O = [Al(H_2O_4)_4(OH)_2]^{1+} + H_3O^+$$

C Hydrolysis products undergo polymerization reactions such as the following:



D The complex most important in coagulation is a polynuclear complex containing 6 or 8 aluminum ions carrying a tripositive or tetrapositive charge.

$$[Al_6(OH)_{15}]^{3+} \quad or \quad [Al_8(OH)_{20}]^{4+}$$

X FACTORS INFLUENCING COAGULATION

A Optimum coagulation represents the attainment of a very complex equilibrium in which many variables are involved.

B Effect of pH. Single most important variable.

1 There is at least one pH zone for any given water in which coagulation occurs in the shortest time, with the minimum coagulant dose and yields an effluent containing the least amount of residual coagulant.

2 pH zone of good coagulation coincides with the zone of least solubility for the hydrolysis products of the coagulant.

3 For aluminum salts pH zone is 6.0 to 7.8, for iron salts pH zone is 4.5 to 11.0.

C Effect of Salts

1 Principal effects of certain ions are to alter

a pH range of optimum coagulation

b Time for flocculation

c Optimum coagulant dose

d Residual coagulant in the effluent

2 Cations have little or no effect on coagulation

3 Monovalent anions have little effect on coagulation

4 Di- and tri-valent anions affect coagulation by:

a Broadening pH zone of coagulation, for example $SO_4^{=}$

b Shifting the pH zone of coagulation, for example $PO_4^{\equiv}$, shift to more acid pH's

D Nature of Particles

1 In general, coagulation is controlled more by the coagulant and chemical composition of the water than by the nature of particles.

2 A certain minimum of coagulant must be added for any suspension to provide enmeshing mass of floc.

3 Dosage of coagulant does not increase linearly with increase of turbidity.

4 Paradoxically, very high turbidities are easier to coagulate than very low turbidities.

5 Hydrophobic particles are easier to coagulate than hydrophilic.

E Effect of Coagulant

1 Alum not widely used.

2 Iron salts have broader pH range for good coagulation.

3 Iron salts preferable for color removal at acid pH's and for coagulation at softening pH's.

F Effect of Physical Factors

1 Temperature is only physical factor to consider.

2 Temperature affects viscosity of water and rate of chemical reactions.

3 Very low temperatures, < 5°C, adversely affects coagulation - requiring more coagulant and longer time.

XI MIXING

A Two stages of mixing or other means to create turbulence are generally used.

B A rapid mixing to uniformly mix the coagulant and promote contact with particles. Thirty to 60 seconds are usually provided.

C A slower and longer stirring to obtain flocculation by floc growth. Detention times as little as 10 minutes but more frequently 30-60 minutes are used.

XII FLOCCULATION

A The rate of flocculation is proportional to the mean velocity gradient (G), the detention time, the concentration of particles and the particle size.

$$G = \left[\frac{C_D A v^3}{2 \nu V}\right]^{\frac{1}{2}}$$ where

G = mean velocity gradient - sec^{-1}

C_D = drag coefficient - for flat blades made of wood C_D is taken to be 1.8, for metal blades 1.2

A = area of paddles - ft^2

v = mean velocity of the paddles relative to the water, ft/sec, generally taken as $V_W = 0.75\ V_P$

ν = kinematic viscosity of the water - ft^2/sec

V = volume of flocculator - ft^3

C Time for flocculation can be expressed by product of "G" and "t" which yields a dimensionless value Gt.

1 Gt values may range within the limits of 10^4 and 10^5

D Design Criteria

1 Rapid mix compartment

a Detention time - 30-60 sec

b Design "G" - 500 sec^{-1}

c Gt - 30,000

2 **Flocculation** chamber

a **Detention** time - 10-30 min

b **Design** "G" - 20-30 sec^{-1}

c Gt - 54,000

Gt values up to 200,000 have been suggested to produce a dense sludge.

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BASIC PRINCIPLES OF CARBON ADSORPTION

I INTRODUCTION

Adsorption can be described as the condensation of gases, liquids, or dissolved substances on the surfaces of solids. Almost any material has some adsorptive capacity, but the effects are not evident unless the adsorbing material (adsorber) is porous and has a very large area for a given mass.

II FACTORS TO BE CONSIDERED

A Contact Time

Adequate time must be allowed to obtain the desired removal.

1 Adsorption usually consists of a rapid initial rate of uptake followed by a marked decrease in rate. The latter rate can be interpreted as intraparticle diffusion and the uptake varies almost proportionately with the half-power of time. Linear variation of the quantity adsorbed with $t^{1/2}$ can be expected.

B pH

The effect of pH is significant in adsorption. An inverse relationship between capacity and pH has been observed. Acidic materials adsorb best at low pH. A free acid adsorbs better than a salt of that same acid.

C Temperature

Temperature effects on adsorption equilibria generally are not felt to be significant, particularly over the range of temperatures encountered in water and wastewater. Increasing temperature should result in decreasing capacity because physical adsorption is ordinarily an exothermic process, however, if viscosity is a factor, the benefits of a decrease in viscosity may override other temperature effects.

D Concentration of Solute

The mechanism, as previously stated, probably includes rapid formation of an equilibrium concentration at the liquid-particle interface followed by slow diffusion into the pores of the particle. A simple relationship between adsorption rate and the solution concentration cannot be expected. A further consideration is the proportion of contaminants that are adsorbable in the solution. Higher concentrations should give more rapid initial adsorption, but this has no relation to the amount of contaminant adsorbed at equilibrium. Tests using a single solute indicate the adsorbed film obtained at very high concentrations of solute is different from that obtained in the lower concentrations common to wastewaters. Low concentration equilibrium data correlate best with a Langmuir equation. Relatively high concentrations are best described by a Brunauer-Emmett-Teller equation.

E Molecular Size of Solute Molecules

If the diffusion of solute molecules is a rate-determining step in adsorption, the size of the diffusing molecule should be a factor. The larger the molecule, the lower the rate of diffusion, and, therefore, the rate of adsorption. This has proven to be the case.

F Viscosity

A highly viscous solvent would decrease the rate of diffusion.

G Carbon Particle Size

A decrease in carbon particle size should give an increase in rate and quantity of adsorption. More surface area is exposed and more pores should be opened for diffusion.

H Carbon Concentration

Data from experiments using the high carbon concentration (powdered carbon) necessary for wastewater treatment are not available. It is expected that the carbon concentration would not affect the rate of adsorption, except for a slight decrease in the rate due to the decreased concentration of the solute.

An increase in the amount of carbon would, of course, increase the quantity of contaminant adsorbed.

I Carbon Pore Structure

Large pores (macropores) completely permeate each carbon particle and act as access ways for diffusion of materials. The macropores contribute very little to the total surface area. The micropores are largely responsible for the adsorptive action. These pores are many orders of magnitude smaller than the macropores and can be thought of as tributaries to the macropores. The diameter of the micropores are in the order of 10 to 100 Angstrom units and, although the micropores constitute only one-third to one-half of the volume of the carbon particle, they contribute as much as 1,000 square meters of surface per gram of carbon, representing virtually the entire surface area. A carbon with very fine pores will adsorb small molecules while leaving larger ones in solution. A carbon with large pores will preferentially remove large molecules while the smaller ones are displaced from the surface.

J Active Surface Area

The total surface area that has specific attractive power for the wastewater contaminants and the accessibility of such surface is an important factor. Comparing two carbons, the one with more active, accessible surface area will remove more of the contaminants.

III ADSORPTION THEORY

Various theories have been proposed for the phenomenon of adsorption, but due to the many factors involved the concept is very complex. No single mathematical equation other than a purely thermodynamic one can cover all cases.

Adsorption is termed reversible if the same equilibrium conditions are reached when the solute concentration is increased to a definite level from a lower concentration as when it is decreased from a higher concentration to the same level. At equilibrium, the rate of adsorption is just equal to the rate of desorption. Reversibility is observed with many solutes although it is not a universal phenomenon; many adsorptions are not reversible or are incompletely reversible.

The equilibrium relation between the concentration of a substance in the adsorbed state and the concentration remaining in solution forms the basis of most equations that have been developed to represent adsorption. Many of these equations are useful in specific cases, but none are universally applicable. It is for this reason that the simplest relation is commonly used.

A Freundlich Adsorption Isotherm

A simple relationship is found by conducting adsorption experiments through a range of equilibrium concentrations and determining the amount of substance adsorbed at each equilibrium concentration.

The mass of material adsorbed is related to the concentration of the solution by the equation

$$\frac{x}{m} = kC^{\frac{1}{n}}$$

where x = units of material adsorbed

m = weight of adsorbent

C = equilibrium concentration of materials remaining unadsorbed in solution k & n = constants which have different values for each solute and adsorbent.

If we take logarithms of both sides of the equation, we get:

$$\log x/m = \log k + \frac{1}{n} \log C$$

If we now plot log x/m against log C we should get a straight line. This expression is usually known as the Freundlich adsorption isotherm. It has no theoretical basis, but is simply empirical. The curve is termed an adsorption isotherm because the curve changes with temperature, but is fixed for one given temperature.

B Langmuir's Formula

Langmuir's formula has a better theoretical basis than that of Freundlich's.

Suppose that, for 1 sq. cm. of a given surface, there are N available sites for adsorption. At equilibrium, let n of these sites be occupied; there will then be (N-n) free sites. As stated before, the rate of adsorption is just equal to the rate of desorption at equilibrium. The rate of adsorption is proportional to the concentration of the solute and to the number of free sites, that is,

$$\text{rate of adsorption} = k_a (N-n)C$$

The symbols x, m, and C have the same definitions as in the Freundlich isotherm. The rate of desorption depends only on the number of occupied places, that is,

$$\text{rate of desorption} = k_d n$$

Thus at equilibrium

$$k_a (N-n)C = k_d n$$

and

$$\frac{k_a NC}{k_d + k_a C} = \frac{k_1 C}{1 + k_2 C}$$

where $k_1 = k_a N/k_d$ and $k_2 = k_a/k_d$. But x/m is directly proportional to the number of occupied places, n per s cm. From which Langmuir's adsorption isotherm follows

$$x/m = \frac{k_1 C}{1 + k_2 C}$$

An important difference between the two theories is that Langmuir's formula takes into account the fact that, for very large values of C, the surface becomes saturated, and can take up no more. For large values of C, k_2C greatly exceeds 1, so the latter can be neglected and the equation then becomes

$$x/m = \frac{k_1}{k_2}$$

that is, the mass of substance adsorbed reaches a constant value which is not exceeded although C may be increased. Also, for small values of C, k_2C may be neglected in comparison with unity. The equation then becomes

$$x/m = k_1 C$$

showing that the amount adsorbed is proportional to the concentration when the surface is nearly bare.

The Langmuir theory assumes that the maximum adsorption corresponds to a single layer of solute molecules on the adsorbent surface.

C Physical-Chemical Adsorption

In many early studies of gas adsorption, it was observed that the more condensable gases are usually adsorbed in larger quantities. Because of this, adsorption was explained as resulting from forces similar to those which produce liquefaction. According to this view, adsorption will occur when the physical attraction of a solid surface is greater than the attraction of the solvent and other solute molecules for the affected molecules. The cohesive physical forces that cause the molecules to be attracted to each other are called van der Waal's forces.

There are adsorptions, however, in which the behavior indicates that a chemical change is involved. Adsorption of this type is termed chemical adsorption or chemisorption.

Physical adsorption usually involves a smaller energy change than does chemisorption. Chemisorption is specific and depends on the chemical nature of both the adsorbent and adsorbate. The selectivity in physical adsorption is usually due to purely physical properties. Physical adsorption decreases at higher temperatures, while higher temperatures are often needed for chemisorption.

IV SUMMARY

In most adsorption systems, a high initial rate of adsorption is followed by a rapid decrease in the rate. Days or weeks may be required to obtain equilibrium. The rate after the initial adsorption appears to be controlled by diffusion processes. The diffusion process has been shown to take place in accordance with the half power of time. If the rate of adsorption is wholly dependent upon the surface area, then it should be inversely proportional to the first power of the diameter, but, for intraparticle diffusion, the variation seems to be with the reciprocal of the square of the diameter of the particle. The intraparticle diffusion, following initial adsorption, seems to be the controlling factor. Many factors affect the adsorption process, temperature, concentration, pH, etc., all being important. Using simple formulae to explain the adsorption of the mixtures of materials found in natural and wastewaters is impossible. Some materials hinder the adsorption of other constituents or may aid in the process. In general, the net effect of mixed materials is not detrimental. Research has shown the following generalities can be made:

1 Weak electrolytes are adsorbed better than strong electrolytes.

2 The more ionized a material is, the more difficult it is to adsorb it out of solution.

3 Slightly soluble materials are generally adsorbed better than highly soluble materials.

4 High-molecular-weight materials may be adsorbed better than those of low molecular weight.

There are, of course, exceptions to any generalities, including those given above.

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BASIC PRINCIPLES OF ION EXCHANGE

I INTRODUCTION

Ion exchange is a well known method for softening water and for producing deionized water. It may also be practical for removing from wastewater the inorganic materials added during use. Removal of this mineral increment is necessary if the quality of the water is to be preserved. It may also be applicable to the removal of nutrient materials that are becoming increasingly serious as a pollution problem.

II MECHANISM OF ION EXCHANGE

Ion exchange materials are those containing ions that can be replaced by other ions from solution. They may be liquids, but are usually solids with a particle size range that makes them convenient for use in a column. A typical solid ion exchange material can be looked upon as a microporous matrix containing many ionic sites - either all positive or all negative - as shown in Figure 1. To maintain electroneutrality each ionic site must have associated with it an ion of opposite sign (counter ion). This counter ion must be mobile enough so that it may be replaced by, or exchanged for, another ion of the same sign upon contact with a solution of this new ion. When all the original counter ion is replaced, the ion exchange material is exhausted. A concentrated solution of the original counter ion must be used to regenerate the ion exchanger. If the counter ion is a cation, the ion exchange material is a cation exchanger. If the counter ion is an anion, the ion exchange material is an anion exchanger. All the common cations and anions in wastewater may take part in ion exchange. Hydrogen and hydroxyl ion are also often involved.

III MATERIALS HAVING ION EXCHANGE PROPERTIES

Proper design of ion exchange systems requires considerable knowledge concerning the physical and chemical nature of exchange materials. Common types of ion exchange materials include:

A Minerals

The most common of these are the zeolites. They are crystalline aluminosilicates with cation exchange properties. They have been used often to soften water by exchanging sodium ion for calcium and magnesium.

B Coals

These contain carboxylic and possibly other groups that give cation exchange properties. Chemical treatment, especially by sulfonation, improves the chemical and physical properties.

C Synthetic Organic Resins

Since their development, these resins have become the most important of all ion exchange materials. They can be divided generally into the following four groups:

1 Strong acid resins

These are characterized by containing strongly ionized anionic groups, such as sulfonate, in the organic matrix. A typical structure is shown in Figure 2. Since the counter ions are cations, the resins are cation exchangers. They have the ability of "salt splitting" or exchanging hydrogen ion for the cation of a salt such as NaCl. A typical reaction is

$$RSO_3^-H^+ + Na^+ \longrightarrow RSO_3^-Na^+ + H^+$$

2 Weak acid resins

These are cation resins containing groups that are only partly ionized, such as carboxyl. They are not effective for salt splitting, but in the free acid form will remove cations from basic

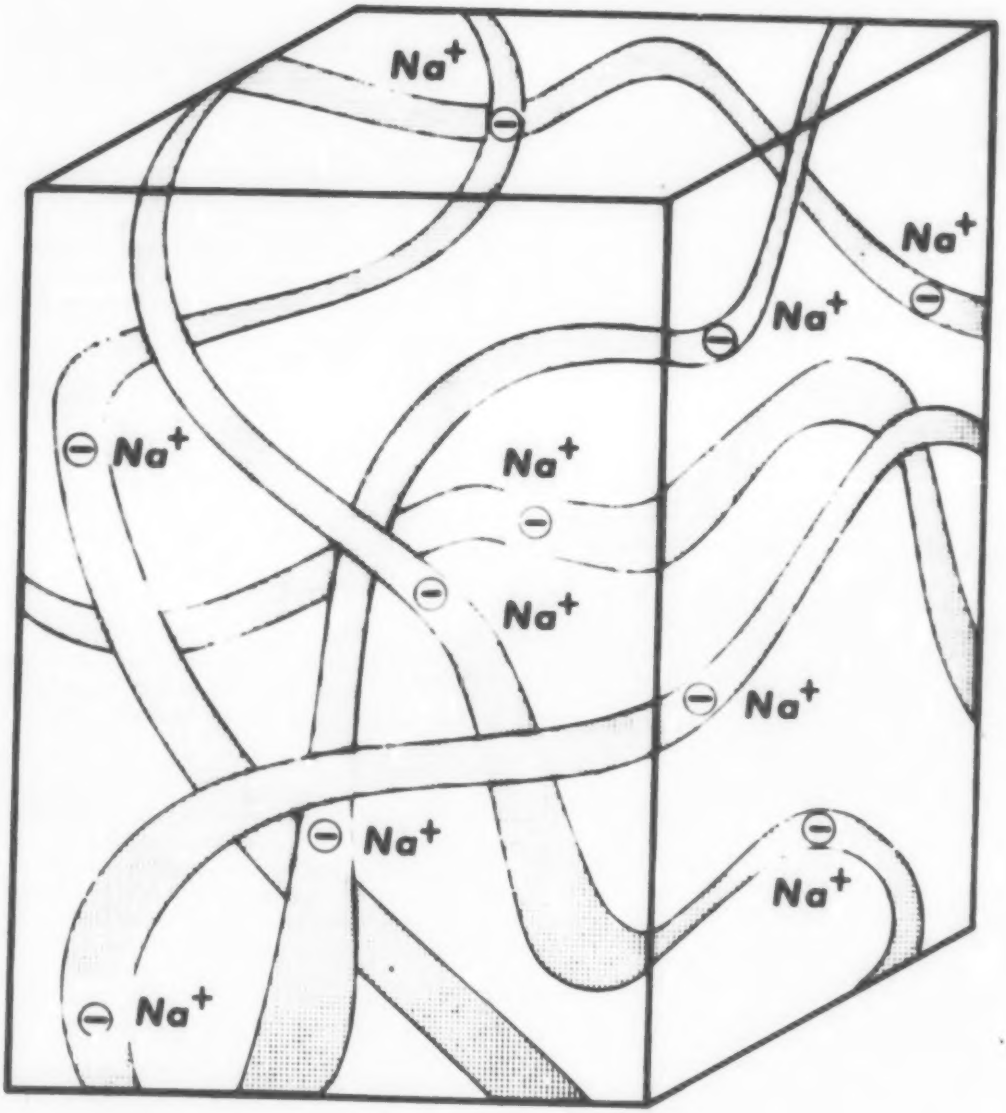
Na+
Na+
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FIGURE I.
TYPICAL EXCHANGE MATERIAL

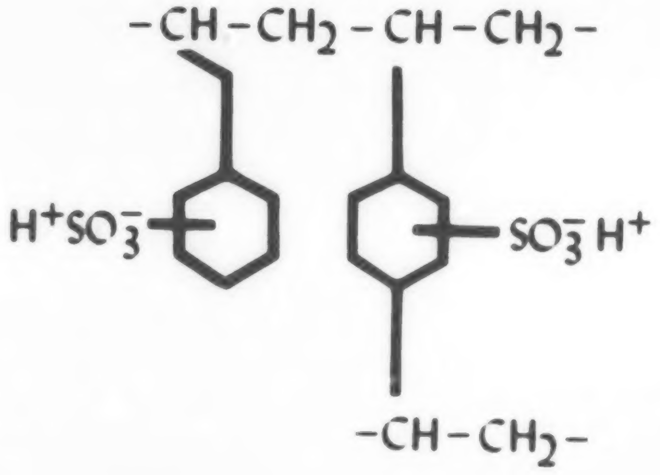


FIGURE 2. STRONG ACID RESIN

solutions. A typical reaction is

$$RCOOH + NaOH \longrightarrow RCOO^-Na^+ + H_2O$$

3 Strong base resins

These anion exchange resins are often quaternary ammonium compounds. They are strongly ionized and can accomplish salt splitting. A typical reaction is

$$RN(CH_3)_3{}^+OH^- + Cl^- \longrightarrow RN(CH_3)_3{}^+Cl^- + OH^-$$

4 Weak base resins

These anion resins are not highly ionized in the hydroxyl or free base form. They are actually acid absorbers and are not usually considered to be salt splitting. A typical reaction is

$$RNH_2 + HCl \longrightarrow RNH_3{}^+Cl^-$$

D Liquid Ion Exchangers

Ion exchange can take place between two immiscible liquids. Liquids that are immiscible with water, may for example, function as ion exchangers if they contain ionogenic compounds. Materials are available that can be dissolved in organic solvents to give cation or anion exchangers.

IV SELECTIVITY

A Non Specific Exchangers

Although a general-purpose cation exchange material will remove any of the common cations from a wastewater, there is competition between the various cations. Similarly, there is competition between the various anions for the anion exchanger. If a sample of resin is allowed to remain in a solution containing two counter ions, at equilibrium there may be a much larger amount of one counter ion on the resin than the other. There are a number of rules of thumb for determining what ion will be preferred. Two that are likely to apply to wastewater state that the counter ion of higher valence and the counter ion of smaller solvated equivalent volume will be preferred. For cations the order of preference for some common examples is $Ca^{++} > K^{+} > NH_4^{+} > Na^{+}$. For strong acid resins, hydrogen ion would be to the right of sodium. Weak acid resins have strong affinity for hydrogen ion. For anions the order of preference for two examples is $SO_4^{--} > Cl^{-}$. For strong base resins, hydroxyl would be to the right of Cl^{-}. Weak base resins have strong affinity for hydroxyl.

B Exchangers with Specific Selectivity

There are applications for ion exchangers highly specific for certain ions. In domestic wastewater treatment, there is presently a demand for methods to remove nutrients. The most important of these are nitrate, phosphate, and ammonium ion. Of the three, only an ammonia selective material is presently being investigated. Resins specific for the other nutrient ions could be of value.

V EXCHANGE CAPACITY

A Total Capacity

Ion exchange materials have a total exchange capacity determined by the number of active sites on the matrix. The following discussion pertains primarily to the solid synthetic organic resins. The total capacity is usually expressed by manufacturers as meq/dry gram or meq/ml where volume is bed volume, not resin particle volume. The latter method of expressing capacity is convenient for the engineer who must deal with total bed volume. Care must be used to define the counter ion associated with the ion exchange material ("form" of the resin) when giving capacity. On a weight basis the different equivalent weight of different counter ions will obviously affect resin weight. On a volume basis, the effect of the counter ion is even greater since different counter ions swell the resin to different degrees.

B Useful Capacity

The total capacity of ion exchangers is never used completely in practice. Consider a column of cation exchange resins having hydrogen as the counter ion. A sodium chloride solution passed through the column will exchange sodium for hydrogen. Because of diffusion resistance in the liquid and in the resin particles, exchange is not instantaneous. If it were, complete exhaustion of each resin particle would be possible. Instead, the resin will be exhausted as shown in Figure 3. As exhaustion proceeds, some sodium will begin to breakthrough in the product water. The ion exchange resin must then be regenerated with an acid solution. As Figure 3 shows, the resin capacity will not all be used by the time regeneration is required because the resin at the end of the column has not become completely exhausted. The useful capacity is further decreased after regeneration. Regeneration would not be continued until all the counter ions were replaced by hydrogen. That resin remaining in the sodium form would be lost for further sodium removal.

Utilization of resin capacity can usually be improved by increasing the column length or by changing conditions so that diffusion rate is increased.

VI REGENERATION

A Cocurrent

In the usual ion exchange installation, the ion exchange resin is contained in a column that is fed downward. Upon exhaustion, the bed is backwashed and regenerated in a downward direction. This cocurrent method of regeneration makes poor use of both resin capacity and regenerant. In Figure 3, for example, passing acid in the same direction as the sodium chloride

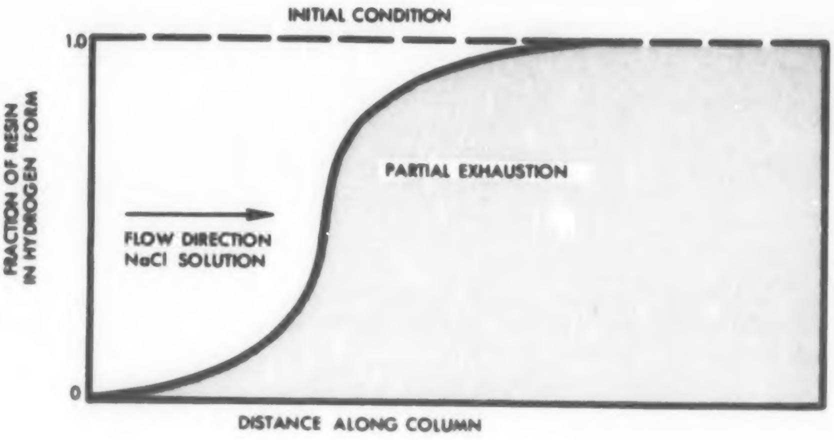


Figure 3. Effect of passing NaCl Solution through Cation Resin in the Hydrogen Form.

solution results in the movement of sodium into the resin that had not been completely exhausted. A large excess of acid is required to remove the sodium. Unless essentially all the sodium is removed it will contribute to early breakthrough during the next ion exchange cycle. This is shown in the data of Table 1 which were taken by a manufacturer using a common cation resin.

Table 1

Regenerant Level (meg/ml resin)	Useful Capacity (meg/ml resin)
0.76	0.48
4.56	1.47
7.6	1.60

Data from Rohm & Hass, "Technical Notes, AMBERLITE IR-120" revised March 1960.

Sodium chloride was fed under the same conditions for each run and the breakthrough criterion was the same. Acid was then used for regeneration. Obviously, a compromise must be made between capacity and amount of regenerant. Economics determine the regenerant level chosen.

B Countercurrent Regeneration

Although operating problems have prevented the extensive use of countercurrent regeneration, this method may be a necessity to make ion exchange competitive with other processes that are being developed for treatment of wastewater. In countercurrent regeneration the least exhausted resin contacts the most concentrated regenerant. The driving force for regeneration of this resin is greater than with cocurrent regeneration. The overall driving force for the whole column is also increased. During ion exchange the resin at the outlet of the column is highly regenerated so that early breakthrough is prevented.

This outline was prepared by Dr. Carl A. Brunner, Research Chemical Engineer, Contracts and Grants Activities, Advanced Waste Treatment Research Laboratory, NERC, EPA, Cincinnati, OH 45268.

ELECTRODIALYSIS

I INTRODUCTION

Usually when water is used, there are added varying amounts of inorganic materials. In the case of domestic wastes, the increment amounts to about 300 to 400 mg/l. Most of the inorganic material is composed of ions common to natural waters. Also added are phosphate and nitrogen as ammonia or nitrate. Removal of the added material is obviously necessary to maintain the quality of the water. With industrial wastes, especially from industries dealing with metals, the inorganic materials added are heavy metal ions that may be very toxic. Their removal is even more important than the removal of the ions usually found in natural waters.

Electrodialysis is useful for the partial demineralization of fairly dilute waters. For practical application usually the total dissolved solids (TDS) concentration is about 2000 mg/l or less. Since municipal wastewaters nearly always have a TDS less than 1000 mg/l they are well within the range where the process should be economically feasible. Many industrial wastes also undoubtedly fall into the practical concentration range. With electrodialysis sufficient demineralization is usually carried out to reduce the TDS to 500 mg/l or slightly less. Further demineralization results in appreciable added cost.

II PRINCIPLES

A direct electric voltage impressed across a cell containing mineralized water will cause positively charged ions or cations to migrate to the negative electrode and negatively charged ions or anions to migrate to the positive electrode. If cation- and anion-permeable membranes are placed alternately between the electrodes as shown in Figure 1, alternate compartments become more concentrated while the intervening compartments become diluted. Many membranes can be placed between the electrodes forming many dilute and many concentrate compartments. Manifolds can then be added so that the membrane stack can be fed continuously and demineralized product and waste concentrate streams can be removed continuously.

The cation(+) and anion(-) permeable membranes that make electrodialysis possible are composed of ion exchange materials. Cation membranes are likely to be sulfonated styrene with divinylbenzene added for cross linking and strengthening of the polymer. Anion membranes are also often styrene base with various amine groups being used to give the ion exchange property. The sulfonate and amine groups make the membranes very hydrophilic. Membranes absorb appreciable water when in contact with aqueous solutions.

Although the structure and function of the membranes is complicated, a rough quantitative explanation of the reason for their ability to transport ions of one charge can be obtained from consideration of the Donnan principle. This principle, in an incompletely rigorous form, states that the ion concentration product of a compound will be equal for water in the membrane and for the surrounding solution. A cation membrane in contact with very dilute sodium chloride can be considered as an example. The product of the concentrations of the sodium and chloride ions will be very small in dilute solution. In the structure of the membrane there will be a high concentration of fixed negative charges which, to maintain electroneutrality, will attract an equal number of positive charges or in this case sodium ions. The concentration of sodium ions associated with the water in the membrane will then be very high in relation to the original solution concentration. By the Donnan principle the chloride ion concentration in the membrane must be very small. If a voltage is applied across the membrane, sodium and chloride ions will move in opposite directions through the membrane. Because of the preponderance of sodium ions, however, most of the current will be carried by that ion. The cation

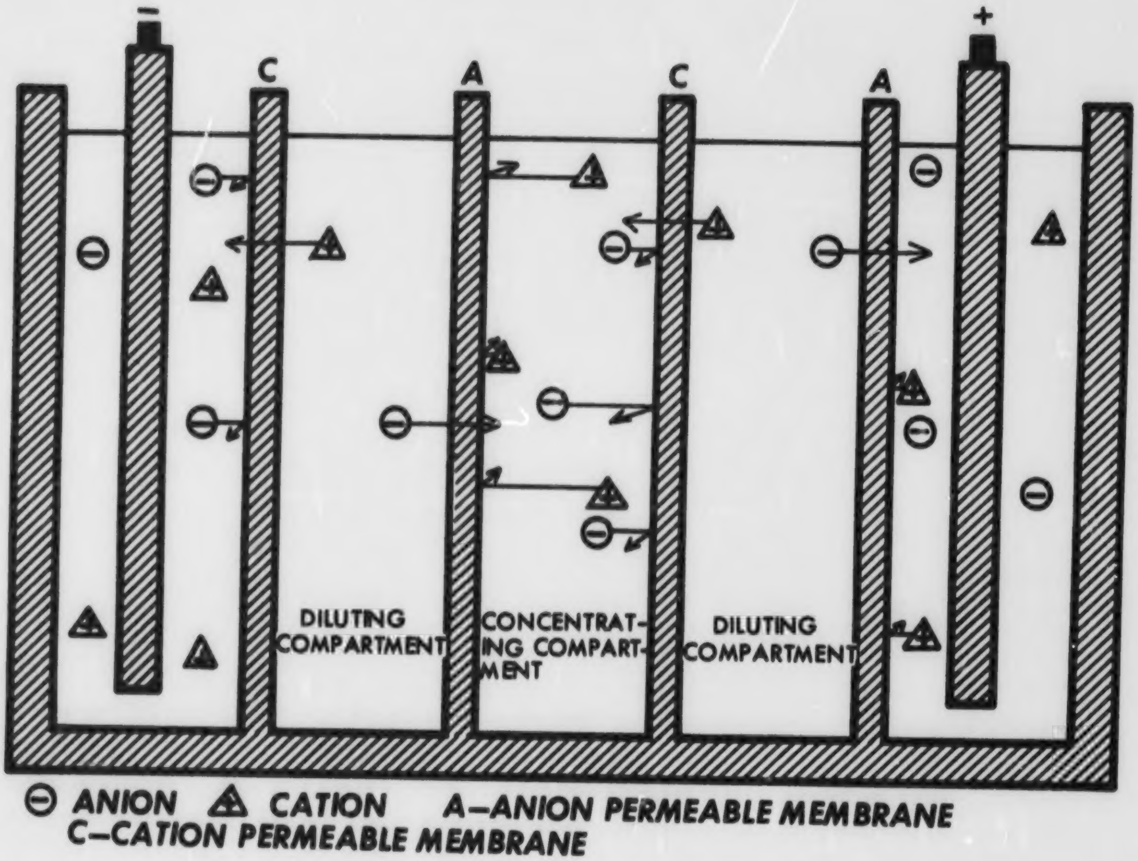

−
+
C
A
C
A
DILUTING COMPARTMENT
CONCENTRATING COMPARTMENT
DILUTING COMPARTMENT
⊖ ANION
△ CATION
A–ANION PERMEABLE MEMBRANE
C–CATION PERMEABLE MEMBRANE

Figure 1. Electrodialysis Principle

membrane will, therefore, have a high selectivity for sodium and for cations in general. A similar argument can be used to show why anion membranes are selective to anions. As the Donnan principle indicates, selectivity decreases as solution concentration increases.

One of the important factors to be considered in operating an electrodialysis stack is the electrical power requirements. In treatment of wastewater, fortunately, this is not as important as in demineralization of a more highly mineralized water. Since direct current is used for electrodialysis, the stack power is simply the product of the current and voltage. The demineralizing current can be calculated from the product of feed rate per diluting compartment, concentration change from feed to product water expressed as normality, and Faraday's constant, 96,500 coulombs/equivalent. The required voltage is more difficult to determine. Classical thermodynamics allows the minimum voltage to be calculated (1) for a membrane separating solutions of different concentrations. In the usual electrodialysis application this minimum voltage, which only applies strictly at a zero current, is a small fraction of the total. The usual approach is to consider the stack as a series of resistances for which the voltage can be calculated from Ohm's law. Resistances that should be included are diluting and concentrating liquid streams, membranes, and the thin stagnant liquid films adjacent to the membranes. Usually, however, the resistances other than those of liquid streams are simply combined as a single value which must be determined experimentally. This empirical resistance can then be used to estimate the approximate voltage for an electrodialysis stack using the same kind of membranes. To calculate the total power requirements the voltage requirements of the electrodes and electrode compartment streams must be added to the voltage for the remainder of the stack.

III OPERATING PROBLEMS

The stagnant liquid films that occur at the membranes are responsible for significant problems during operation. Serious concentration gradients can develop across these films because of the higher electrical transference numbers in the membranes compared to transference numbers in solution. As a result there is a decrease in concentration from bulk dilute stream to the membrane surface and an increase in concentration from bulk concentrate stream to the membrane surface. The high concentration at the face of the membranes on the concentrating side leads to scaling from precipitation of compounds with low solubility. The low mineral ion concentration at the dilute side of the membranes causes transfer of hydrogen and hydroxyl ions from the water through the membranes. Polarization is the term used to describe operation when water decomposition becomes significant. The decomposition of water into ions wastes electric power. A more important effect, however, is the increased scaling potential at the concentrate side of the anion membranes. Hydroxyl ion increases the precipitation of magnesium hydroxide and calcium carbonate.

An effect that is common to electrodialysis is membrane fouling. This phenomenon is different from scaling since it involves deposition of materials on the dilute side of the membranes. Although there can be a number of causes in wastewater treatment fouling appears to be due to electrophoretic movement of negatively charged colloidal particles to the anion membranes. Because of the size of the particles they cannot move through the membrane. Instead they form a layer at the membrane surface that interferes with demineralization.

IV APPLICATION TO MUNICIPAL WASTEWATERS

A Laboratory Investigation

Bench scale study (2) of electrodialysis for treatment of municipal secondary effluent indicated that good removal of soluble and insoluble organic materials was necessary for satisfactory operation. Cellulose cartridge filters were used for removal of the suspended organic solids. These filters, however, would not be practical on a large scale and would have to be replaced by another form of suspended solids removal. Granular activated carbon was used for soluble organic removal. Using adequate pretreatment the cost of electrodialysis, exclusive of pretreatment and waste concentrate disposal, was estimated at less than 10 cents/1000 gal.

B Pilot Plant Investigation

As a result of encouraging results from the laboratory study a pilot scale investigation was undertaken (3). Equipment was installed at the Lebanon, Ohio Sewage Treatment Plant. Secondary effluent was the feed to the system. Based upon the experience gained from the laboratory work, clarification and soluble organic removal were considered necessary pretreatment for the electrodialysis stack feed. Diatomaceous earth filtration, using for the most part an alum treated water grade filter aid, was chosen as the method of clarification and granular carbon in fixed beds was chosen for adsorption of soluble organics. The pilot system is shown in Figure 2. The electrodialysis stack is an Ionics Mark III capable of holding up to 150 cell pairs. It is designed to remove about 40 percent of the TDS from the feed. Membranes for this stack are 18 × 40 in. and have about 3000 m^2 active area. Spacer material and edge gasket are combined in one piece. Spacer thickness is 0.032 in. Using all 150 pairs the nominal capacity of the stack is 50 gpm product water. As a result of difficulties with the diatomaceous earth filter, it was necessary to reduce the feed water rate. The number of cell pairs was reduced to 125 and the product rate to 42 gpm. The concentrate waste rate at reduced feed rate has been 4 gpm. Internal recirculation of concentrate at a rate about equal to the product rate is necessary, however, to maintain approximately equal pressures on both sides of the membranes. The pH of the concentrate stream is held at 5 or less to prevent calcium carbonate scale formation.

Operation of the pilot plant has resulted in only one serious problem, fouling of anion membranes. Membrane fouling increases stack resistance and necessitates higher voltage to maintain the desired degree of demineralization. Figure 3 shows how demineralization at constant voltage was affected by fouling. Fouling rate depends to some extent upon feed turbidity as might be expected. The unexpectedly low fouling rate for the run made after the stack was acid rinsed at a low enough pH to kill organisms suggests biological growth on the membranes as a contributing factor. Fortunately fouling has not been permanent. Shutdown over a weekend has always restored the fraction demineralization to near normal. Even with some fouling, operation is considered practical since excess stack capacity can be provided to take care of down time for self-cleaning of membranes. Control of feed turbidity is necessary for practical run lengths before shutdown is required.

Scaling of anion membranes, a problem that is serious in many electrodialysis installations, has not been particularly troublesome in treatment of municipal wastewater. Calcium carbonate scale formation has occurred on occasions, but has not done serious damage to membranes. Failure to control the pH to 5 or less can cause serious scaling that ruins the anion membranes. In the pilot work concentrate blowdown has been held at 9.5 percent of the product. Scaling potential would be reduced if the amount of blowdown were increased. Such an increase would, however, increase the problem of brine disposal. From a pollution control standpoint, therefore, the concentrated waste must be kept as low in volume as possible.

Permanent damage to membranes during pilot runs over a period of 1500 hours has not been great. Membrane pair resistance at the beginning of runs, for example, only increased about 5 percent even though fouling conditions during some runs were severe. Current efficiency, that is the fraction of the current that results in demineralization, has not changed significantly. Serious membrane damage would result in decreased current efficiency.

To obtain better control of turbidity and more dependable operation than was possible with the diatomaceous earth filter, a chemical clarification system has recently been installed before the carbon columns. This system consists of an upflow clarifier and dual media filters. Lime has been chosen for use in the clarifier because of the softening, phosphate removal, and alkalinity removal that this material offers. This change in pretreatment has not eliminated membrane fouling. The fouling rate is roughly the same as was obtainable with the diatomaceous earth filter when it was producing water of less than 0.1 JTU water.

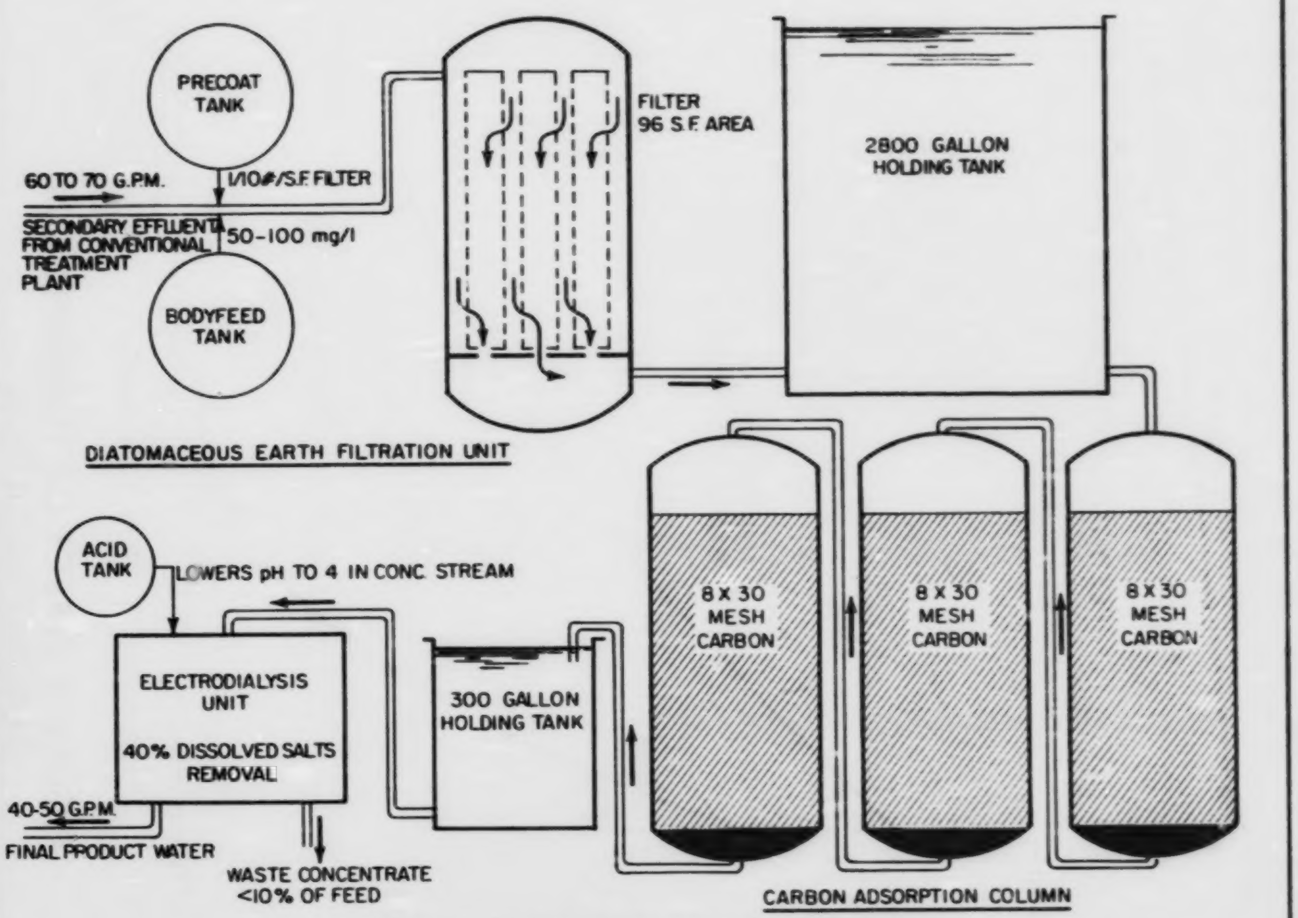

PRECOAT TANK
60 TO 70 G.P.M.
1/10#/S.F. FILTER
SECONDARY EFFLUENT FROM CONVENTIONAL TREATMENT PLANT
50-100 mg/l
BODYFEED TANK
FILTER 96 S.F. AREA
2800 GALLON HOLDING TANK
DIATOMACEOUS EARTH FILTRATION UNIT
ACID TANK
LOWERS pH TO 4 IN CONC. STREAM
ELECTRODIALYSIS UNIT
40% DISSOLVED SALTS REMOVAL
40-50 G.P.M.
FINAL PRODUCT WATER
WASTE CONCENTRATE <10% OF FEED
300 GALLON HOLDING TANK
8 X 30 MESH CARBON
8 X 30 MESH CARBON
8 X 30 MESH CARBON
CARBON ADSORPTION COLUMN

FIGURE 2. ELECTRODIALYSIS PILOT PLANT
WITH DIATOMACEOUS EARTH FILTER PRE-TREATMENT
LEBANON, OHIO

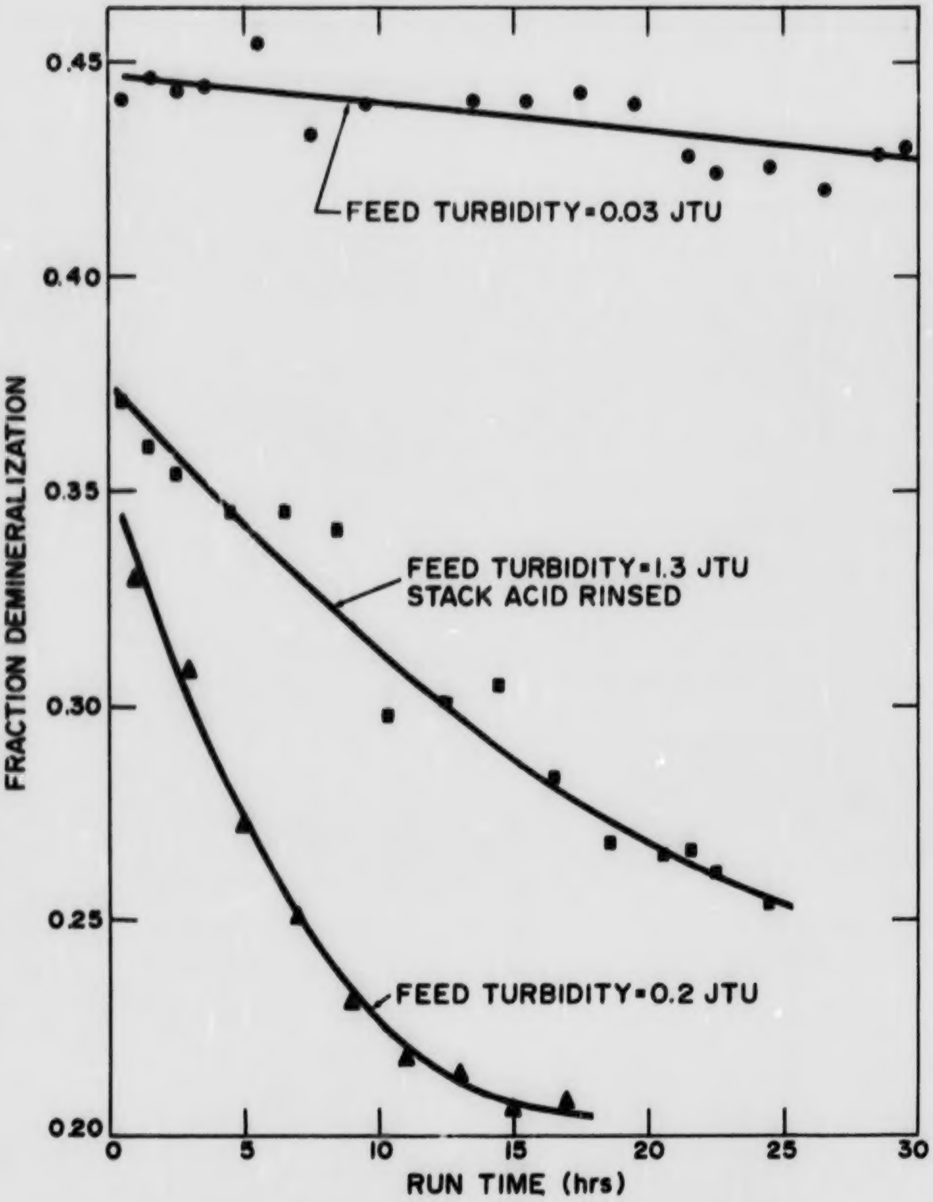


FIGURE 3. EFFECT OF TURBIDITY ON DEMINERALIZATION .

Further work with chemical clarification will be carried out.

C Other Investigations

Ionics Incorporated has operated a small pilot electrodialysis stack at Orange County on trickling filter effluent that had been alum clarified, chlorinated, and treated with granular activated carbon. In addition the stack had before it 10 and 3 micron cartridge filters and an ultraviolet light for sterilization. The fouling rate apparently was considerably lower than observed during operation of the Lebanon stack. Operation was halted after more than 800 hours of run time when a serious stack resistance increases occurred.

More recently this same electrodialysis system, including cartridge filters and ultraviolet light, has been operated on carbon treated, chlorinated secondary effluent at Pomona California. Here carbon treatment is carried out without any clarification. Frequent replacement of the cartridge filters is necessary. Preliminary results suggest that fouling will be at a lower rate than observed on the Lebanon stack. Further work is required to verify the early results and to determine the effect of the cartridge filters and ultraviolet light.

V APPLICATION TO BRACKISH WATER

The principal use for electrodialysis has been in treatment of brackish waters. These are inland waters containing enough salts to prevent or limit direct use. The mineral content is considerably less than for sea water, often about 2000 mg/l, and their compositions can vary widely from one location to another. Electrodialysis has been found practical for demineralizing these waters to 500 mg/l TDS or slightly less.

There are problems associated with brackish water treatment, some of which are different from those observed in wastewater treatment. Fouling is sometimes troublesome, and is often not of organic origin. Instead inorganic materials such as traces of iron and manganese can cause the difficulty. It is usually considered necessary to remove essentially all the iron and manganese from ground water to be treated by electrodialysis. Scaling is often a problem because of high concentration of materials of low solubility such as calcium sulfate and calcium carbonate. To avoid scale formation the amount of concentrate waste is kept considerably higher than is possible in wastewater treatment. Often the volume of concentrated waste is from 33 to 100 percent of the volume of treated water.

There is an increasing number of electrodialysis installations using brackish water in the United States. Two of these are particularly well known. They are at Webster, South Dakota and Buckeye, Arizona. The Webster plant began operation in 1962 as a Demonstration Plant by the Office of Saline water and has a nominal capacity of 250,000 gpd. Electrodialysis equipment was supplied by the Asahi Chemical Industry Company of Japan. The feed water is not unusually high in TDS for brackish water since it contains only about 1700 mg/l. It is, however, a very difficult water to treat. It is high in calcium, sulfate, and bicarbonate so that scale formation is a serious problem. The water also contains significant amounts of iron and manganese that must be removed by using potassium permanganate injection and filtration through manganese zeolites before electrodialysis treatment. One of the wells that was to provide part of the feed water contains organic materials that cause a rapid increase in stack resistance. Many problems have arisen, therefore, during operation of this plant. These have been described in detail in a number of Office of Saline Water publications and summarized in a 1965 paper (4). As a result of the problems the Webster site has been used to test a number of pretreatment methods, both for iron and manganese removal and softening. An interesting feature of the plant as now operated is the use of periodic polarity reversal to reduce scale formation. Improved operation is obtained at a slight loss in current efficiency. Successful use of polarity reversal could represent a significant breakthrough in the application of electrodialysis to high-hardness, badly-scaling waters.

Results at Buckeye, Arizona have been more optimistic than results at Webster. The water, although higher in TDS, has a relatively large amount of sodium and chloride ions. Scaling is, therefore, much less likely to be a problem. The water contains no manganese and little iron. The equipment installed at Buckeye was supplied by Ionics, Inc. and has a design capacity of 650,000 gpd. Over the period October 1962, through June 1965, it is reported (5) that there were no major malfunctions. During that time the average load factor was only 32 percent. This unexpectedly low figure resulted partly from the recirculation of water through evaporative coolers. Recirculation was not possible with the more highly mineralized brackish water available before installation of the electrodialysis equipment and, therefore, the daily water consumption was higher. The low load factor has resulted in a higher than expected water cost when amortization of equipment is included, but the installation is, nevertheless, considered a success.

VI INDUSTRIAL APPLICATIONS

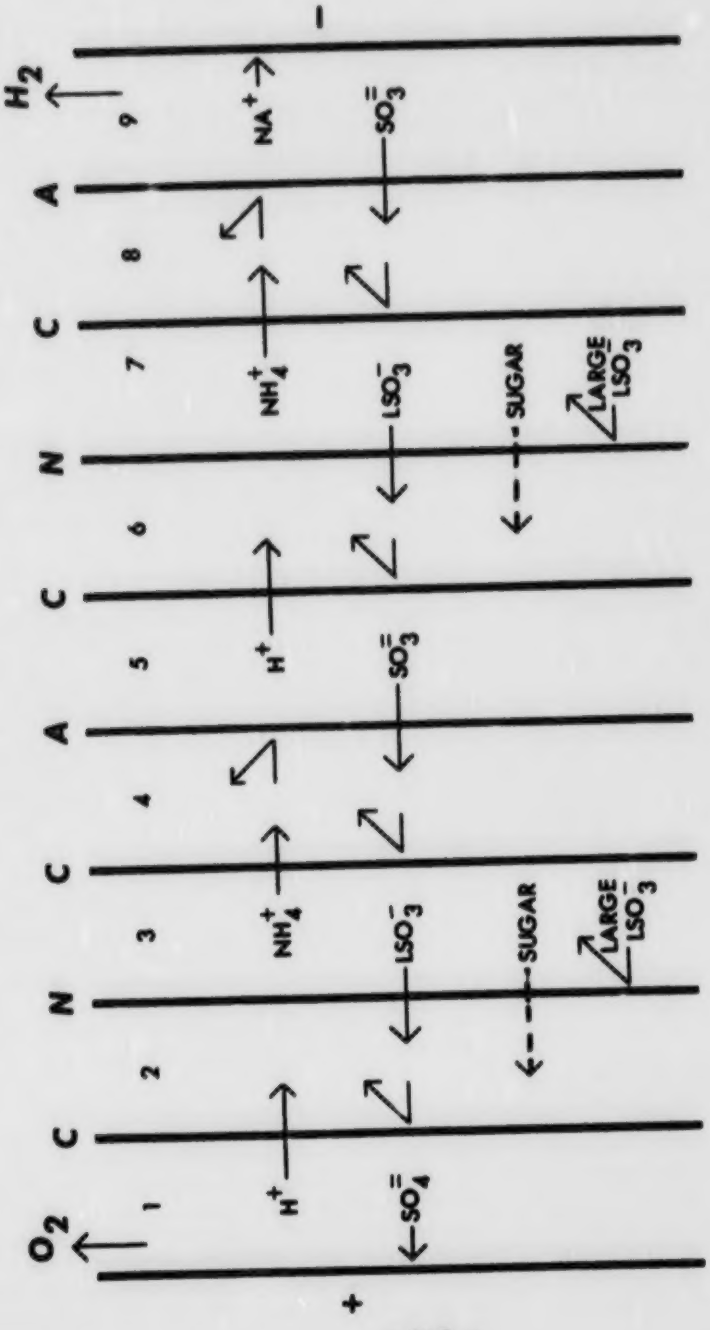
Use of electrodialysis for treatment of industrial wastes has been very limited. Two possible applications that have a pollution control aspect are treatment of pickle liquor and treatment of spent sulfite liquor. The first application is actually more nearly electrolysis than electrodialysis and results in recovery of iron and sulfuric acid from the waste (6). There is more interest in this process in Europe than in the United States. The second process was developed by the Sulfite Pulp Manufacturers Research League (7) and results in recovery of cations from the spent liquor. The principle of this modified electrodialysis process is shown in Figure 4. Spent liquor enters compartments 3 and 7 where cations and organics such as small ligno-sulfonate ions are removed. New liquor is made up in compartments 4 and 8. Compartments 5 and 9 serve as sources for sulfite ion. The spent liquor with low molecular weight organics removed may have a number of uses. The process operates at high current efficiencies and high current densities. Because of the arrangement of membranes, the anion membranes are protected from fouling by organics in the spent liquor.

Although electrodialysis has not yet been used extensively for treatment of industrial wastes, it may have wider application in the future. Its practicability is improved if useful materials can be recovered from the treatment. With the present cost picture for electrodialysis, recovery is probably a necessity. It is accomplished in the two examples cited above. There are undoubtedly many industrial waste streams where chemical recovery is a possibility.

In its basic form of alternating dilute and concentrate streams the process would ordinarily only be applicable to streams free of fouling soluble organics and suspended materials. By using special membrane arrangements such as in the sulfite liquor treatment, fouling problems can be reduced. It should be recognized that in applications where useful materials are recovered the added maintenance resulting from membrane fouling may not be prohibitive in cost.

The maximum concentration that can be economically treated would be limited to the brackish water range if by-product recovery were not possible. Electric power consumption for removing a large concentration of ions, ordinarily becomes prohibitive. Where useful materials are recovered the maximum concentration could be increased until the cost for electrodialysis exceeds the cost of other processes such as distillation. A detailed study of each particular application would have to be made to determine whether electrodialysis is practical.

There are two limitations to electrodialysis that may prevent its use in some applications. One limitation is the inability to practically remove ions below a concentration of several hundred mg/l. Where waste are very dilute, ion exchange should be more appropriate. The other limitation is the inability for the process to be very selective for any one material in a mixture. At present a valuable metal ion could not in general be removed


O2
H2
+
-
C
N
C
A
C
N
C
A
1
2
3
4
5
6
7
8
9
H+
SO4=
NH4+
LSO3-
SUGAR
LARGE LSO3-
H+
SO3=
NA+
SO3=

C—CATION SELECTIVE MEMBRANE
A—ANION SELECTIVE MEMBRANE
N—NON—SELECTIVE MEMBRANE

Figure 4. Electrodialysis Process for Treating Spent Sulfite Liquor

101<

with high selectivity from common ions such as sodium and calcium. Relatively little work has been done to develop selective systems, however, and improvements are likely if a strong need arises. Again ion exchange may be a more practical form of treatment.

VII ULTIMATE DISPOSAL OF CONCENTRATED WASTE

Experience in the use of electrodialysis on municipal wastewater indicates that the volume of the concentrated waste will be 5 to 10 percent of the product volume. To prevent pollution of surface waters by this highly mineralized waste it must either be transported to the sea or in some other way be treated for disposal. Disposal, no matter what the method, will not be insignificant in cost except near the sea. For brackish water the potential disposal problem is also great. Presently the treatment of brackish water is limited enough that disposal of brine has not been a serious problem. Industrial waste treatment would not prevent a significant disposal problem if recovery of by-products from the concentrated waste were practiced. In that case the waste would be further treated anyway. If recovery were not possible, disposal or treatment of the concentrate to make it innocuous would be necessary. The volume of concentrate would vary widely depending upon the particular waste. If it is low in membrane scaling materials, the volume may be only a few percent of the product. Each case must be considered separately.

VIII EQUIPMENT DESIGN

Since electrodialysis is a recently developed process, design of equipment is not yet highly sophisticated. The major American manufacturer, for example, has available a limited number of stack configurations to be used primarily for brackish water. Minor modifications are made in the equipment to adapt it to other uses. It would be a fortunate coincidence if obtainable equipment were the optimum for any given application. Because of the limited market, manufacturers have not been able to justify the extensive research and development work that is required for optimized design.

Several discussions of equipment design have been published (8, 9). Application of the methods were also demonstrated for the case of municipal wastewater (2). The reader is referred to these publications for details. In principle, these methods depend upon Ohm's Law and Faraday's Law and material balances. The resulting equations allow membrane area and power requirements to be calculated. Before making design calculations, it is necessary to know membrane resistances and limiting current density. These are empirical quantities, but manufacturers should have the necessary data available for their particular stack configurations and for brackish water feeds. Experimental data might be required for design of equipment to treat an industrial waste.

One serious fault with published design methods is that they do not emphasize the design of membrane spacers. After the designer chooses an available spacer and has the necessary empirical data it is relatively easy to size the equipment and determine power requirements. Optimum design of a spacer is a much more difficult problem. Presently, spacers range in form from rather open types that give low velocity and pressure drop to tortuous path types that give high velocity and pressure drop. To reduce production costs, manufacturers tend to pick one type and use it for all applications. The Office of Saline Water is presently sponsoring work on spacer design. Preliminary results suggest that even the brackish water spacers may be improved significantly. Further work is obviously justified.

At the present time, the user of electrodialysis equipment usually depends heavily upon the limited number of manufacturers for the ultimate design of a system. The number of types of stack components available is small. Use of the equipment is not yet so widespread that consulting engineers are experienced enough to undertake design themselves. If the market increases substantially, the situation will change. Not only will a greater variety of components become available, resulting in more flexibility of design, but the designing itself will be done by a greater number of people.

IX ELECTRODIALYSIS COSTS

A Capital Costs

When discussing the capital cost of electrodialysis, distinction must be made between present and projected costs. The market for this equipment has not expanded to the point where mass production of components is possible. The costs of membranes and spacers especially are very high. Capital costs do not represent, therefore, the minimum that would be possible just from an increased market and the competition that should develop.

It is difficult to make capital cost estimates without consulting a manufacturer. There is not sufficient information available in published form to make the precise estimates possible. Ionics, Inc. has at various times made available in company literature plant costs for their equipment as a function of capacity and feed concentration assuming the feed to be brackish water and product to contain 500 mg/l TDS. Table 1 shows some typical costs. These figures do not include pretreatment beyond a protective cartridge filter. Cost is strongly affected by both feed concentration and capacity. An independent cost estimate has been made for treatment of secondary effluent based partly upon experience from pilot operation. For a 10-mgd plant and using Ionics equipment, the estimated installed cost is $0.34/gpd. This is significantly higher than the Ionics estimate for 900 mg/l brackish water to which it should most closely compare. Much of the difference can be accounted for in the conservative feed rate per cell pair that was chosen for the wastewater estimate. A small amount was contributed by the inclusion of pH control equipment on the concentrate streams. This equipment is not usually supplied for brackish water treatment, but is desirable on wastewater treatment where there can be frequent changes in alkalinity.

The Office of Saline Water has, in the past, supported work on a procedure for estimating the cost of electrodialysis applied to brackish water. More recently, work was done on optimization of the process. This work will appear in publications from that organization.

B Operating Cost

Long term operating costs for electro-

Table 1

Investment Cost for Electrodialysis Plants

Plant Capacity (mgd)	Feed Concentration (mg/l)	Installed Plant Cost ($/gpd)
2	900	0.33
2	3,000	0.69
10	900	0.19
10	3,000	0.39
100	900	0.11
100	3,000	0.25

dialysis are available from only a few brackish water installations. Estimates are usually based upon experience combined with assumptions about long term maintenance requirements. Membrane life is probably the most questionable factor. Usually a 20 percent replacement per year is assumed. Pilot experience on wastewater suggests that membrane life is greatly affected by the care that is taken in handling the membranes and in operating the equipment. Although definite proof is not yet available, there is reason to believe that an annual 20 percent replacement may not be necessary. Ionics has estimated the cost of electrodialysis operation over a wide range of capacities and feed water concentrations. Some examples are given in Table 2. The estimates are for brackish water and assume 90 percent load factor and 20 percent annual membrane replacement. They include amortization of equipment. An independent estimate for the treatment of 10 mgd of wastewater was 16¢/1,000 gal. This is somewhat higher than the 12¢/1,000 gal for the nearly comparable 900 mg/l brackish water. The difference can be accounted for in the conservatively high capital cost estimate for the wastewater plant.

The effect of load factor on operating cost is significant. At the Buckeye, Arizona, brackish water plant, for example, it is estimated that at full load, the total cost of producing water should be 32¢/1,000 gal. The plant was actually designed to operate at a 48 percent load factor and produce water for 52¢/1,000 gal. For a number of reasons the load factor declined below the design value. For the first 32 months of operation it was only 32 percent. The result was a water cost of about 70¢/1,000 gal. This is an extreme example, but it points out the need to consider ways for minimizing fluctuations in flow. For industrial wastes, it may be possible to make extensive use of relatively cheap water storage to allow operation at a high load factor.

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Table 2

Operating Cost for Electrodialysis

Plant Capacity (mgd)	Feed Concentration (mg/l)	Total Operating Cost (¢/1,000 gal)
2	900	16
2	3,000	39
10	900	12
10	3,000	29
100	900	8
100	3,000	21

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This outline was prepared by Dr. Carl A. Brunner, Research Chemical Engineer, Contracts and Grants Activities, Advanced Waste Treatment Research Laboratory, RATWRC, NERC, EPA, Cincinnati, OH 45268.

REVERSE OSMOSIS

I INTRODUCTION

A The phenomenon of osmosis was first observed in the mid-18th century when studies were being made on biological membranes.

B It was noted that when two solutions of different concentration were separated by a semi-permeable membrane, the solvent would flow from the dilute to the concentrated side. A semi-permeable membrane is defined as one that will permit passage of some materials (usually a solvent) while rejecting others.

C The first experiments with artifically-prepared membranes were conducted in the mid-19th century when a membrane was made by precipitating copper ferrocyanide in the pores of porcelain.

D By 1920, interest in osmosis waned and, except for its biological importance, it was considered a laboratory curiosity.

E Developments in the last ten years by C. E. Reid and others of the University of Florida and S. Loeb and others at UCLA showed that the process can be reversed by applying pressure to the higher concentration side and that it has potential as a process for recovering reuseable water from a contaminated or saline source.

II THEORY

A Figure 1 illustrates the principle of osmosis and reverse osmosis. At equilibrium, the pressure on the more concentrated solution is known as the osmotic pressure. This pressure is dependent entirely on the difference in concentration between the two solutions and is not a function of the type of membrane, provided it is semi-permeable.

B It is well known that the "activity" or "chemical potential" of a solvent decreases when in a solution. Thus, the activity of water in distilled water is greater than the activity of water in a salt solution. This greater activity is thought to be the driving force that, in normal osmosis, causes water to flow through the membrane. In brackish and sea water, this driving force is about 1 psi for every 100 ppm salt in solution.

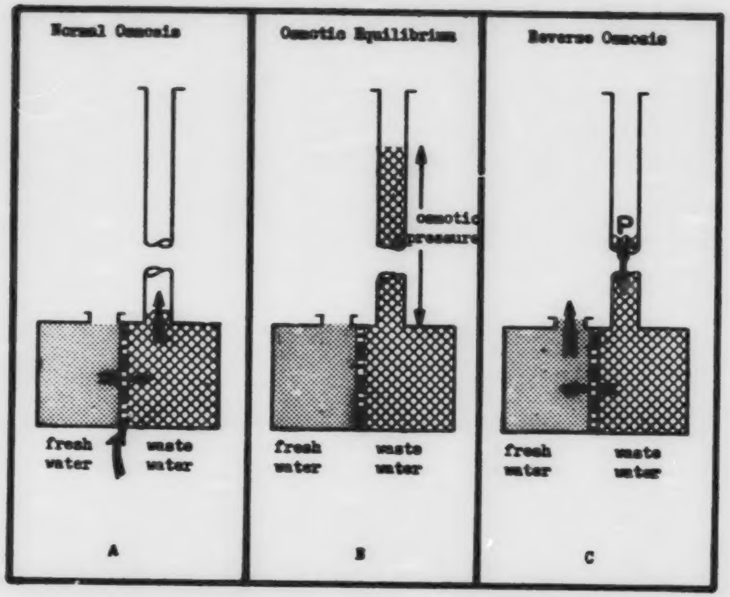
C The actual mechanism of transfer of water through a reverse osmosis membrane is not completely understood. The most generally-accepted theory is that water passes through the membrane by successive transfer from one adsorption site to the next. In the case of cellulose acetate, the most common membrane material, the adsorption forces arise from hydrogen bonding.

D The rate of transfer of water through a membrane in reverse osmosis is directly proportional to the applied pressure minus the osmotic (back) pressure and inversely proportional to the thickness of the membrane. Increasing temperature, since it raises the activity of the water, will result in a greater water flux through the membrane.

E As stated before, the flux of water is directly proportional to the pressure driving force. The flux of salt, however, is a function only of the salt concentration. Raising the applied pressure, therefore, reduces the salt content of the product.

III MEMBRANES AND THEIR MANUFACTURE

A Reverse osmosis was first made practical by the discovery (by Dr. Loeb and coworkers at UCLA) of a process for making a cellulose acetate membrane with a very thin active layer. These membrane are about 100 microns in total thickness. The active, dense layer that transmits the


Normal Osmosis
Osmotic Equilibrium
Reverse Osmosis
osmotic pressure
P
fresh water
waste water
fresh water
waste water
fresh water
waste water
A
B
C

THE PRINCIPLE OF REVERSE OSMOSIS
(WASTE WATER)
Figure 1

315-63-77

107<.

water and rejects the salt is only 0.25 microns thick - the rest is a spongy back-up material that has little, if any, desalination properties.

B A typical formulation for making cellulose acetate membranes is shown in Table 1.

TABLE 1

CASTING SOLUTION FOR CELLULOSE ACETATE REVERSE OSMOSIS MEMBRANES

Material	Weight
Cellulose Acetate	22.2
Magnesium Perchlorate	1.1
Acetone	66.7
Water	10.0

This casting solution is poured on a glass plate and its thickness is adjusted by a doctor blade. The plate is then immersed in ice water for 30 to 60 minutes during which essentially all of the magnesium perchlorate is removed from the membranes. The membrane is then immersed in hot water for 3-10 minutes. The water temperature is critical - from 55° C on up, the salt rejection properties are improved at the expense of flux through the membrane. Continuous casting procedures have been developed but they are proprietary and nothing has been published on them.

C The degree of acetylation of the cellulose is an important factor. The optimum membrane is produced from cellulose acetate containing about 2.5 acetate groups per cellulose molecule. The tri-acetate is produced and it is hydrolyzed to the 2.5 level. Further hydrolysis degrades the membrane - this is controlled in service by maintaining the pH of the feed water at about 5.5.

D Compression of the spongy under-layer at the elevated pressures necessary during operation results in flux decline. Efforts are underway to produce a thin (0.25 micron) dense layer that could be put on a porous, non-compressible substructure.

E Considerable effort has been and is being expended on the development of new membrane materials. As yet, none of those investigated has proven as effective as cellulose acetate. Some of the more promising materials studied to date include poly (vinylene carbonate), polyvinylpyrollidone and graphitic oxide.

F A recent development in this field has been made at the Oak Ridge National Laboratories of the Atomic Energy Commission. Dr. Kraus and co-workers have developed a dynamically-formed membrane having permeation rates of 100-300 gallons/day ft^2 (gfd) compared to a typical rate for cellulose acetate of 20-30 gfd.

1 The rejecting layers are formed by pumping water containing a hydrous oxide (zinconium or thorium) or a polyelectrolyte [poly (vinylbenzyl-trimethyl-ammonium chloride)] over a porous body.

2 The porous body can have a pore size up to 5 microns in diameter.

3 Salts containing polyvalent counter-ions (Mg^{+++} or $SO_4^=$) are poorly rejected and destroy the rejection capacity of the membranes for mono-valent salts.

IV ENGINEERING OF REVERSE OSMOSIS UNITS

A Several different designs have been proposed to meet the design objective of producing the maximum amount of quality product per unit cost.

1 The "flat plate" membrane configuration is similar to a plate and frame filter press. These units have been tested extensively on brackish- and sea-water but little work has been done on wastewaters.

2 The spiral-wound membrane is illustrated in Figures 2 and 3. The objective of this design is to increase the surface area to volume ratio. This configuration has been evaluated on brackish- and sea-water and is being evaluated on wastewater at the FWPCA Pomona, California pilot plant.

3 The tubular unit is shown in Figure 4. Another form this can take is that of a metal tube with weep-holes drilled for delivery of the product water.

4 The newest configuration consists of hollow fibers with water passing from the outside to the inside of the fibers. Figure 5 shows a schematic of such a unit.

B Two different concepts have been employed in the design of these units. The first three described above depend on relatively high permeation rates (20 gfd). The hollow fiber units rely more on very high surface area per unit volume with low permeation rates (0.15 gfd).

C Equipment must be designed to minimize the "boundary layer" effect - that is, the existence of a high concentration layer at the membrane surface. This problem is related to the permeation rate, becoming more severe as the rate increases.

D Rapid membrane replacement and inexpensive pressure construction are also goals of the designers in this field. Membrane life is still an unknown quantity, particularly in wastewater treatment, so configurations in which membrane replacement is costly could be impractical.

V LABORATORY INVESTIGATIONS ON WASTEWATER

A Aerojet-General conducted a 1-year study on the application of reverse osmosis to wastewater. The equipment consisted of two 0.5 ft^2 membranes in series (flat-plate). The concentrate stream was returned to the feed tank to determine the effect of increasing concentration. The principal findings are listed below.

1 The product quality was excellent. Table 2 shows the product quality from two test runs.

2 Water flux through the membrane decreased rapidly. Acid addition to the feed to maintain a pH of 5 minimized this problem but flux reduction with time was still severe.

3 After a period of several hours, the flux at 1500 psi feed pressure was about the same as that obtained at 750 psi.

4 The flux through "loose" membranes, though high originally, rapidly declined to the same level as the "tight" (low flux, high salt rejection) membranes.

5 Removal of organics from the feed by carbon adsorption was helpful in reducing the rate of flux decline.

6 Test results showed that biological attack of the membrane could be a severe problem.

B The General Atomic Division of General Dynamics ran several spiral-wound modules on wastewater at the Pomona Water Renovation Plant of the Los Angeles County Sanitation Districts. This was a continuous flow-through operation recovering about 5% of the feed as product.

1 Water quality was again excellent as can be seen in Table 3. The "A" modules contained "tight" membranes, the B - "loose" membranes and the "C" - intermediate.

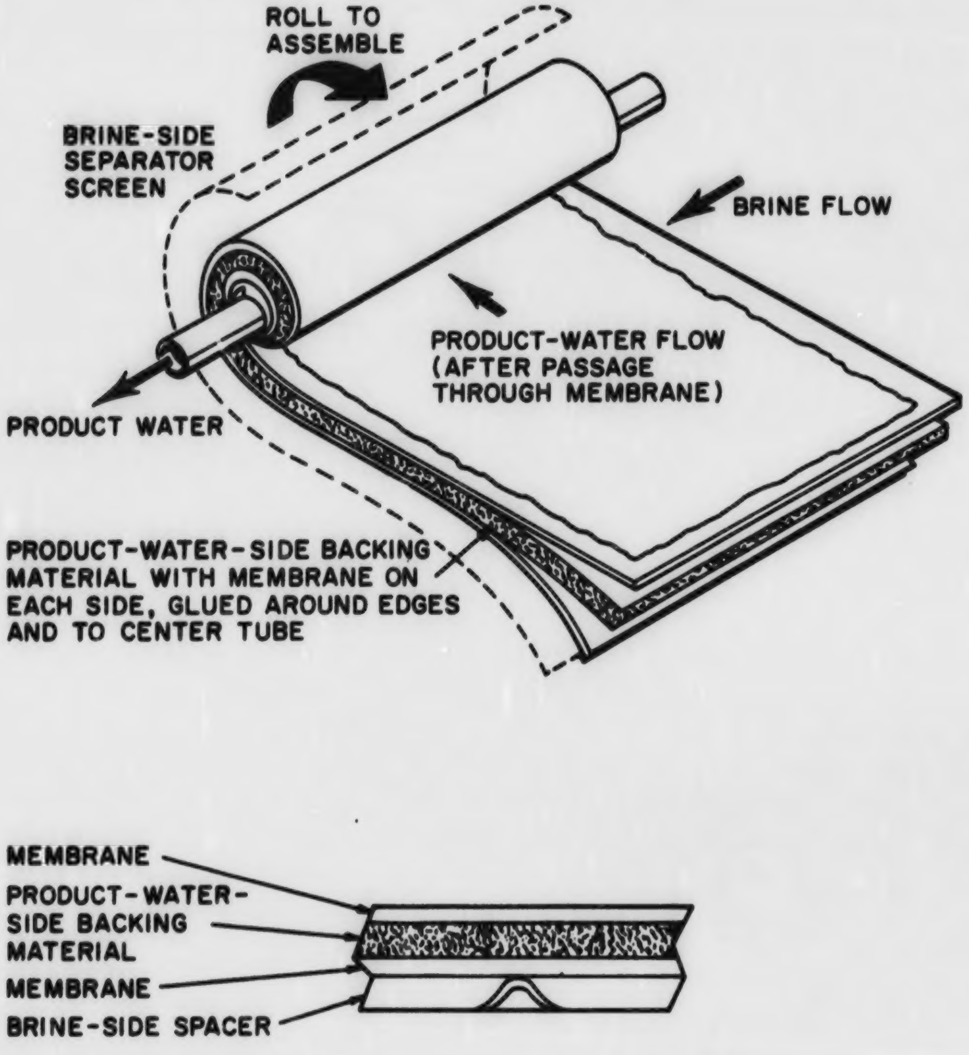

ROLL TO ASSEMBLE
BRINE-SIDE SEPARATOR SCREEN
BRINE FLOW
PRODUCT-WATER FLOW (AFTER PASSAGE THROUGH MEMBRANE)
PRODUCT WATER
PRODUCT-WATER-SIDE BACKING MATERIAL WITH MEMBRANE ON EACH SIDE, GLUED AROUND EDGES AND TO CENTER TUBE
MEMBRANE
PRODUCT-WATER-SIDE BACKING MATERIAL
MEMBRANE
BRINE-SIDE SPACER

Figure 2. A Spiral-Wound Reverse Osmosis Module

Reprinted with permission from General Dynamics,General Atomic Division

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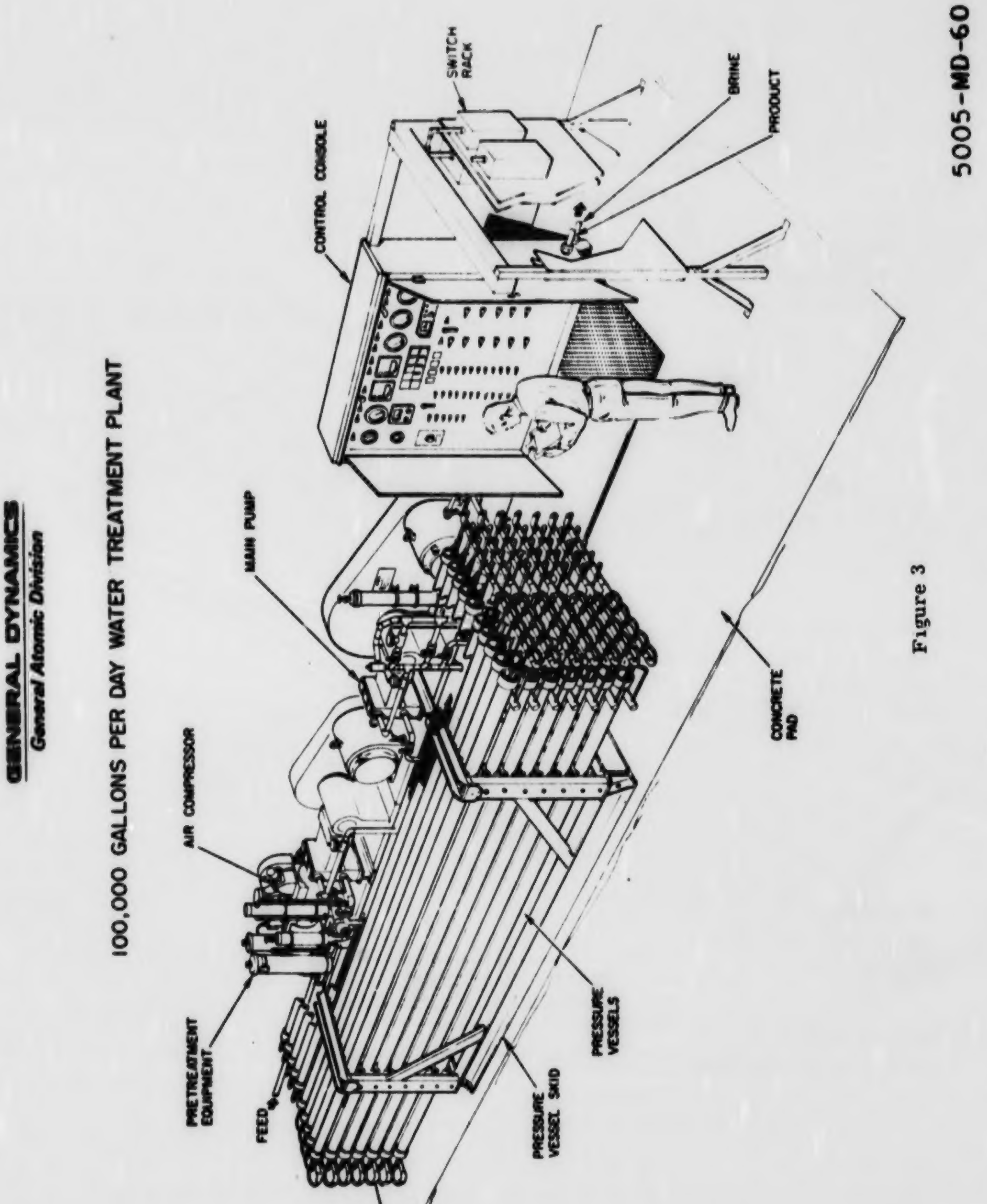
GENERAL DYNAMICS
General Atomic Division
100,000 GALLONS PER DAY WATER TREATMENT PLANT
SWITCH RACK
BRINE
PRODUCT
CONTROL CONSOLE
MAIN PUMP
AIR COMPRESSOR
PRETREATMENT EQUIPMENT
FEED
PRESSURE VESSELS
PRESSURE VESSEL SKID
CONCRETE PAD
5005-MD-60

Figure 3

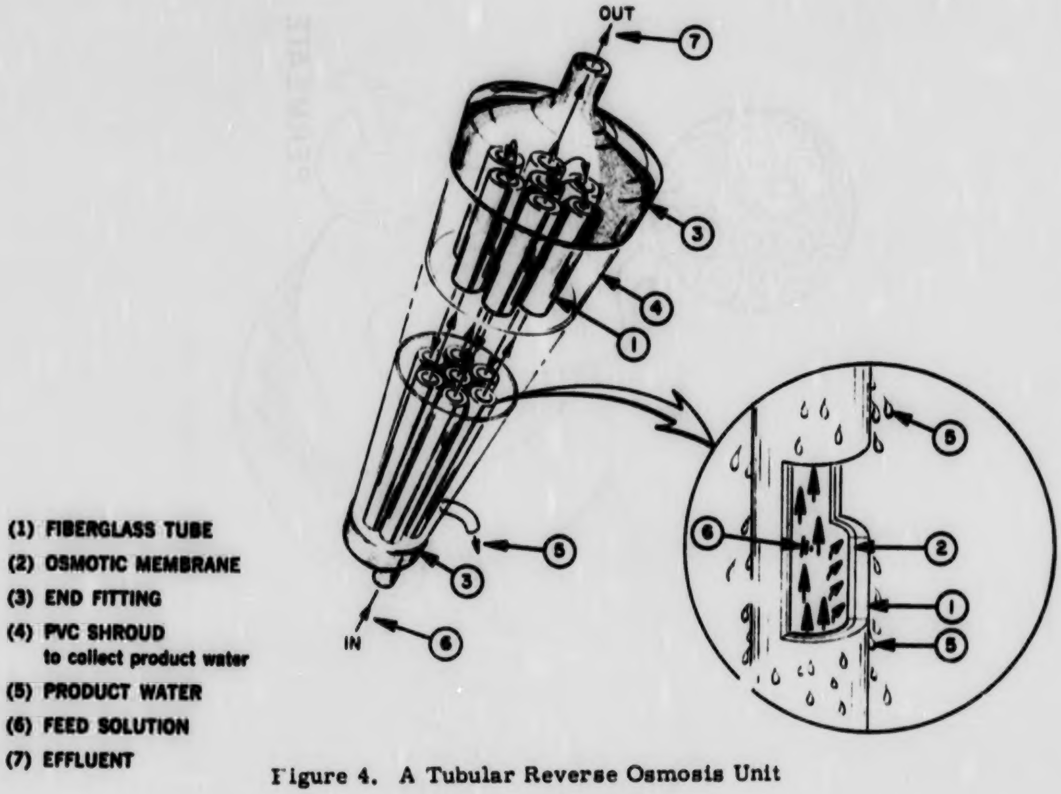
OUT
7
3
4
1
5
2
1
5
6
5
3
IN
6
(1) FIBERGLASS TUBE
(2) OSMOTIC MEMBRANE
(3) END FITTING
(4) PVC SHROUD to collect product water
(5) PRODUCT WATER
(6) FEED SOLUTION
(7) EFFLUENT

Figure 4. A Tubular Reverse Osmosis Unit

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2 As before, the "loose" (high flux-low rejection) membranes plugged fastest and were soon delivering no more than the "tight" membranes.

3 Flux decrease was still a problem, ranging from 25% in 1500 hours for the tight membranes to 86% in 290 hours for the loose membranes.

C Similar work was done on a spiral unit by the New Jersey Department of Health and others on a sewage plant effluent and on water from the Hackensack River. Results were similar. Bacteriological removals were determined during this study and the MPN reduction/ range was from 80 to 99.9%.

VI PILOT INVESTIGATION ON WASTEWATER

A The only pilot investigation on wastewater of reasonable duration has been conducted at the Pomona pilot plant on a 5000 gpd unit utilizing the spiral membrane configuration.

B The first year of operation was plagued by operating difficulties and no encouraging results were obtained. The more significant problems are listed below.

1 Some of the feed by-passed the membrane module, going between the outside of the module and the pressure casing. This caused a flow decrease through the feed channel, thus accentuating the boundary-layer effect.

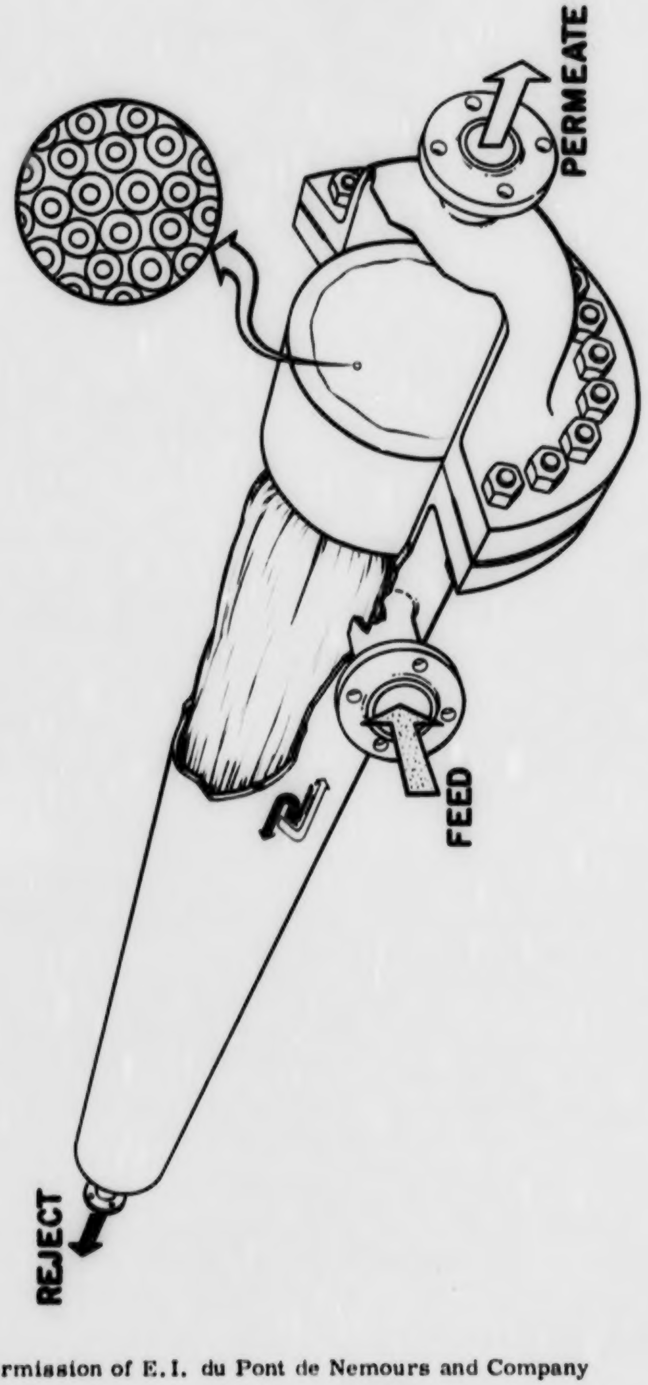
PERMEATE
FEED
REJECT

Figure 5. A Hollow-Fiber Reverse Osmosis Unit

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TABLE 2

WATER MINERAL ANALYSIS

	Feed Water (Typical) ppm	Product (Test 18) ppm	Product (Test 28) ppm
TDS	550	15	28
ABS	4.5	0.1	0.1
COD	95	2	6.0
pH	5^1	6.3	5.5
Cations			
Na^+	85	4.6	6.1
K^{++}	40	4.9	2.7
NH_4^+	25	0	3.9
Ca^{++}	125	0	2.6
Mg^{++}	50	0.7	0
Fe^{++}	N. D.	0	0
Anions			
Cl^-	65	20.3	22.1
NO_3^-	2	0	0
HCO_3^-	260	8.1	3.5
$CO_3^=$	0	0	0
$SO_4^=$	200	1.7	3.3
$SiO_3^=$	30	2.9	2.9
$PO_4^{\equiv}$ (total)	25	0	0

N.D. - not determined

[1] pH adjusted during test

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TABLE 3. PRODUCT QUALITY - REVERSE OSMOSIS TREATMENT OF WASTEWATER

	Feed Water*	Product From Module 1-6-3 5A1 **	Product From Module 3-24-1 3B2 **	Product From Module 1-22-1 5C2 **
Alkalinity, ppm $CaCO_3$	213	< 25	88	37
Ammonia Nitrogen, ppm N	11.2	2.8	7.0	3.5
Total Nitrogen, ppm N	12.6	2.8	7.4	3.5
Specific Conductance, μmhos/cm	933	60	399	215
Chloride, ppm Cl	92	4	68	32
Hardness Total, ppm $CaCO_3$	195	3.8	27.4	6.1
Phosphate Total, ppm PO_4	5.6	0.2	2.2	0.4
Potassium, ppm K	17.8	1.2	10.4	5.0
Sodium, ppm Na	107.2	7.5	63.0	29.5
Calcium, ppm $CaCO_3$	138	2.8	20.3	3.8
Magnesium, ppm $CaCO_3$	58	1.4	7.1	2.3
Dissolved Solids, ppm	531	28	224	118
Sulfate, ppm SO_4	71.4	5.3	2.1	3.9
Fe, ppm Fe	0.10	0	0	0
ABS, ppm	3.8	< 0.1	1.1	<0.1
COD, ppm O	41	4	8	4
pH	7.2	6.6	7.1	6.7

*Chlorinated and diatomaceous earth filtered secondary effluent.

**Pressure = 205 psi
Temperature = 70° F
Time in Use = 1464 hours for 1-6-3 (5A1)
160 hours for 3-24-1 (3B2)
975 hours for 1-22-1 (5C2)

2 The modules tended to "telescope" in the direction of flow due to pressure drop through the feed channels.

3 pH control was erratic and the control set-up was such that the feed pump would shut down when pH control was lost. The pressure surge caused by this sudden loss of pressure accentuated the telescoping and ripped apart some of the taped joints holding the module together.

4 Solids build-up in the feed channels raised the pressure drop through the unit. Calcium sulfate and biological solids were the principal sources of trouble.

C From late June until late September of 1967, the unit has been operating satisfactorily. The feed water was carbon-treated secondary effluent rather than untreated secondary effluent as before.

In addition, pH control was switched from sulfuric to hydrochloric acid and 5 ppm chlorine was added to the feed. The results have been most satisfactory, as listed below.

1 We are operating at a feed rate of about 4800 gallons per day and recovering 83% of the feed as product. The flux rate is 5 gfd at a pressure of 400 psi and there has been no significant decline in flux in 2000 hours of operation.

2 A special anti-telescoping device installed by General Atomic has been successful in overcoming this problem.

3 Product quality has remained high as shown by Table 4. Table 5 shows gradual degradation of product quality between start-up and after 2000 hours on-stream. These increases are probably due to slight pin-hole leaks developing in the membranes.

4 Daily flushing with an air-tap water mixture has been successful in removing any particulate material that accumulates or forms in the modules.

D The study program for the future at Pomona calls for: 1) another attempt at feeding secondary effluent instead of carbon-treated water, 2) using sulfuric rather than hydrochloric acid for pH control, and 3) increasing the operating pressure to obtain 90% recovery of feed.

Table 4. WATER QUALITY DATA FROM GENERAL ATOMIC REVERSE OSMOSIS UNIT

	PO_4 (mg l)		COD (mg l)		NH_3-N (mg l)		NO_3-N (mg l)		TDS	
Tube No.	Ave.	Range	Ave.	Range	Ave.	Range	Ave.	Range	Ave.	Range
Feed	30.9	21.6 - 37.8	10.8	6.7 -14.4	9.2	2.5 - 26.0	2.4	0.8 - 5.0	623	530 - 748
1A	0.22	.09 - .53	1.2	0.0 - 2.3	1.3	0.6 - 2.0	0.4	0.0 - 0.8	51	0 - 133
1B	0.16	.05 - .35	1.1	0.0 - 2.6	1.2	0.5 - 2.2	0.5	0.0 - 0.8	36	9 - 85
2C	1.4	1.3 - 1.5	2.4	0.4 - 6.2	2.8	0.7 - 7.3	0.9	0.4 - 1.7	60	-
3C	1.1	.07 - 3.4	1.6	0.0 - 3.6	1.8	1.0 - 2.8	0.7	0.0 - 1.6	70	14 - 173
4C	0.43	.07 - 1.1	1.8	0.6 - 4.0	1.7	0.9 - 2.6	0.9	0.4 - 1.6	67	22 - 120
5C	0.48	.08 - .90	2.0	0.0 - 5.0	1.7	1.1 - 2.6	0.9	0.4 - 1.3	53	35 - 83
6C	0.41	.08 - 1.0	1.6	0.0 - 3.6	1.8	1.0 - 3.1	1.0	0.7 - 1.6	53	6 - 113
7C	0.37	.04 - 1.8	2.3	0.0 - 6.1	2.6	1.3 - 4.6	1.0	0.5 - 2.4	67	34 - 186
8C	0.77	.07 - 1.8	2.4	0.0 - 5.0	3.2	1.8 - 5.3	1.6	0.6 - 2.6	95	36 - 106
Total Product	0.57	.13 - 1.2	1.7	0.0 - 3.4	1.7	0.9 - 3.4	0.8	0.4 - 1.2	73	30 - 113
Brine	177.	73. -240.	43.8	26.6 -60.4	94.	19. -240.	7.5	2.8 -13.7	3402	2861 -3889
% Reduction	98.2		84		82		67		88	

TABLE 5. PRODUCT WATER CONDUCTIVITY - GENERAL ATOMIC REVERSE OSMOSIS UNIT

Tube No.	Initial Cond. (μmhos/cm)	Present Cond. (μmhos/cm)
1A	50	80
1B	50	80
2C	100*	140
3C	70	200
4C	70	110
5C	75	110
6C	85	120
7C	90	220
8C	110	250

Feed Conductivity: Before acidification: 800
After acidification: 950

Brine Conductivity: 4000

*High due to use of several modules with slight leaks

VII COSTS

A Reverse Osmosis is in an early stage of development and, as could be expected in this stage, there is a wide range on estimated costs. Different manufacturers have estimated costs for a 10 mgd plant ranging from $0.25 to over $1.00/1000 gallons. Some of the factors that will determine the cost of a reverse osmosis system for wastewater renovation are listed below:

1 Flux through the membranes is the prime factor affecting costs. Most cost estimates have been based on an optimistic flux of 20-30 gfd.

2 Membrane life is also an important cost consideration. If the membrane life is short, the reverse osmosis unit will have to be designed so that membrane replacement is rapid and inexpensive. This is an approach being taken by some manufacturers.

3 The degree of pre-treatment necessary prior to entering the reverse osmosis cell will influence application of the process. If organic materials have to be removed before the membranes, then reverse osmosis is simply a demineralization process and must compete economically with electrodialysis and ion exchange.

4 The product to waste ratio obtainable with reverse osmosis is an important consideration. If 90% of the feed is recovered as product, the cost of disposing of the concentrate stream by injection into deep wells (providing the geology is appropriate for this) would be about 2¢/1000 gal. of water produced. If only 80% of the feed can be recovered, the cost of disposal will be 5-6¢/1000 gallons.

5 The production rate of membrane materials will affect the cost of the

operation. The raw material costs are almost negligible but the manufacturing costs are high. Membrane consumption must be high enough to support continuous, automated production facilities.

B Despite all these considerations, reverse osmosis could have widespread application in the future in locations where demineralization of wastewater is required for reuse or pollution control purposes. It must be remembered that reverse osmosis is still an "infant" when compared to processes such as distillation and carbon adsorption. The field is wide open and significant developments are occurring rapidly.

VIII FUTURE PLANS

A The FWPCA has plans for continuing support of development work in the reverse osmosis field, both in testing of new membrane materials and in improvements in engineering design.

B A major undertaking under consideration is the funding of work to determine the minimum pretreatment necessary to prepare a raw sewage for treatment by reverse osmosis. It seems most unlikely that conventional primary and secondary treatment provide the ideal conditioning of the wastewater for the membrane process.

C A major study funded jointly by the FWPCA and the Eastern Municipal Water District of Hemet, California is just getting underway. Starting with secondary effluent, several different types of reverse osmosis units will be run in parallel. Pretreatment of the secondary effluent by any combination of clarification, filtration and carbon adsorption will be possible.

IX INDUSTRIAL APPLICATIONS OF REVERSE OSMOSIS

A The Pulp Manufacturers Research League in Appleton, Wisconsin has investigated reverse osmosis as a method for solving both their water pollution and water supply problems. They hope to operate a 50,000 gpd plant that could be evaluated on several waste streams at a number of pulp mills. Laboratory investigations have been limited to short-term batch runs but the results appear promising.

B The Aerojet-General Corporation has received a grant for a study of the application of reverse osmosis to various industrial waste streams at the Odessa, Texas petrochemical complex. A 560 square foot unit will be mounted on a trailer that will also include a clarifier and diatomaceous earth filter to serve as pretreatment units. The objectives of the study will be to investigate the following factors:

1 Reliability of reverse osmosis as an industrial waste treatment process.

2 Sensitivity of the membrane performance to input water quality.

3 Membrane life.

4 Design data for large-scale units.

5 Economics of the process

Table 6 presents characteristics of some of the waste streams that will be fed to the unit.

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3 Loeb, S. and Johnson, J.S. "Fouling Problems Encountered in a Reverse Osmosis Desalination Pilot Plant."

TABLE 6. ODESSA INDUSTRIAL COMPLEX WASTE WATER STREAMS

Stream	A	B	C	D	E	F	Combined Waste
Total Hardness	80-200 (130)	10-100 (40)	0-30 (15)	130-200 (150)	150-320 (200)	0-625	60-1700 (150)
Ca	21	25	0-8	45	109		20-150
Mg	6	3	0-2	3	8		2-30
Fe	2.5	1	3-5 (4)	7	2	1	<1.0-2.5
NH_3	2-15	1-15	40-125 (60)	0-50 (10)	0-13 (3)	4	1-25 (10)
Total Dissolved Solids	2250-2500	700-1150	4400	10,800	1550-1810	2300	3,500-25,000 (5000)
pH	8.5-9.5	8.5-11.5 (9.6)	9-12 (10)	6.5-7.5	6.5-8.0 (7.3)	8.5-12	8.5-10.0 (9.5)
(COD)	2300	770	2440-3715	140	45-80	1500-7500	650-800
Temperature, F	100-120	96-130 (110)	110-135	68-85	70-95	75-110 (85)	85-110 (90)

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This outline was prepared by Arthur N. Masse, Chief, Municipal Treatment Research Program, Advanced Waste Treatment Research Laboratory, RATWRC, NERC, EPA, Cincinnati, OH 45268.

IN-DEPTH FILTRATION

I DEFINITIONS

A In-Depth Filtration

In-depth filtration is the passage of a fluid through a bed of granular media designed to permit the particulates to be removed to penetrate the media surface and yet be retained within the filter. The degree of penetration and solids removal efficiency can be altered by changing the size and gradation of the filter media or the size and character of the particulates to be removed. In the treatment of wastewater the size and character of the particulates or floc can be changed by pretreating the water prior to filtration. Conventional pretreatment includes chemical coagulation, flocculation and sedimentation.

B Coagulation

Coagulation is the destabilization of a colloidal dispersion by reducing the electrostatic and interionic forces tending to keep the particles apart. In the treatment of wastewater the destabilization is accomplished by the addition of an appropriate chemical in an environment of high turbulence achieved by rapid mixing. This turbulence assures good dispersion of the chemical and increases the opportunity for particle to particle contact.

C Flocculation

Flocculation is the growth in particle size achieved by gentle mixing which provides the opportunity for additional particle to particle contact, bridging, adsorption and physical enmeshment.

D Sedimentation

Sedimentation is the separation of a suspension into a clarified fluid and a more concentrated suspension. In the clarification of wastewaters we are primarily concerned with the settling of flocculating particles as opposed to descrete non-flocculating particles. Secondary consideration is given to thickening within the clarifier.

II THEORETICAL ASPECTS OF CHEMICAL COAGULATION, FLOCCULATION, SEDIMENTATION AND FILTRATION*.

A Coagulation

1 Stability of colloidal dispersions

Colloids can be broadly classified as hydrophillic (water-loving) and hydrophobic (water-hating). Hydrophillic colloids attract water to their surface which engenders stability by preventing particle to particle contact. Hydrophobic colloids achieve stability according to to the classical double layer theory advanced by Helmholtz. The hydrophobic colloids acquire an electrical charge by ionization or adsorption of ions. This charge is usually negative for particles in wastewater. This layer of adsorped ions is tightly bound to the particle and attracts another layer of oppositely charged ions which is also tightly bound. A diffuse layer of positive and negative ions of varying composition forms around the particle. The double-layer prevents particle agglomeration. Since the

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*It is beyond the scope of this manual to give a complete treatise on the theoretical aspects of chemical coagulation, sedimentation and filtration. For those who desire a more extensive review of the subject matter an extensive bibliography on chemical coagulation can be found in reference 1.

thickness of the double layer and the surface charge density are dependent upon the concentration and valence of the ions in solution the stability of the suspension can be reduced by adding appropriate ions. Coagulation is achieved by the introduction of ions having a charge opposite to that of the colloid. Coagulation results not because of a neutralization of charges, but rather the thickness of the double layer is reduced so that the cohesive van der Waals forces are operative. Brownian movement causes particle to particle contact but the probability of contact is greatly improved with mixing. The coagulating power of an ion is dependent upon its valency with a bivalent ion 30-60 times as effective as a monovalent ion and a trivalent ion 600-1000 times as effective.

2 Reactions of primary coagulants

a General

Originally the metal ions were thought to be the effective coagulating agents but it is now known that the hydrolysis products of the metal ions are the coagulating agents. The reactions of aluminum and iron hydrolysis products are very complex. The reactions presented in this manual are over-simplified. A thorough treatment of the hydrolytic chemistry of aluminum and iron is given in reference 1.

b Reactions of aluminum salts

1) Alum

Alum is the most widely used coagulant. Aluminum hydroxide can be formed from the reaction of commercial alum and the bicarbonate alkalinity of the water.

$$Al_2(SO_4)_3 \cdot 14H_2O + 6HCO_3 \longrightarrow 2Al(OH)_3 + 3SO_4^{=} + 6CO_2$$

2) Sodium aluminate

Sodium aluminate is another aluminum salt from which aluminum hydroxide can be formed.

$$NaAlO_2 + HCO_3^- + H_2O \longrightarrow Al(OH)_3 + CO_3^{=} + Na^+$$

There appears to be no difference in the coagulating efficiency of the aluminum hydroxide formed from either salt at equivalent final pH. A combination of the two salts can be used to give precise pH control in unbuffered waters.

$$Al_2(SO_4)_3 \cdot 14H_2O + 6\,NaAlO_2 \longrightarrow 8Al(OH)_3 + 3Na_2SO_4 + 2H_2O$$

c Reactions of iron salts[2]

The reactions of iron salts are very similar to those of aluminum salts.

1) Ferrous sulphate (Copperas)

Usually there is insufficient alkalinity in natural waters to react with the ferrous sulfate. Lime is usually added to produce a floc and avoid soluble compounds of iron remaining in the treated water. The chemical reactions of ferrous sulphate and lime depend somewhat upon the order of addition. The following reaction occurs when lime is added first.

$$FeSO_4 + Ca(OH)_2 \longrightarrow Fe(OH)_2 + CaSO_4$$

$$4Fe(OH)_2 + 2H_2O + O_2 \longrightarrow 4Fe(OH)_3$$

2) Ferric chloride

Ferric chloride is extremely corrosive to metal containers and cannot be dry fed because of its hygroscopic characteristics.

$$FeCl_3 + 3H_2O \longrightarrow Fe(OH)_3 + 3H^+ + 3Cl^-$$

d Reactions of lime

Lime is always applied as slaked lime $Ca(OH)_2$. Some of the lime may not be available because of encrustation in precipitates of calcium phosphate, carbonate or sulfate.

$$Ca(OH)_2 + 2HCO_3^- + Ca^{++} \longrightarrow 2CaCO_3 + 2H_2O \; (pH > 9.5)$$

$$Mg^{++} + Ca(OH)_2 \longrightarrow Mg(OH)_2 + Ca^{++} \; (pH > 11)$$

The gelatinous precipitate $Mg(OH)_2$ gives excellent clarification. In waters low in magnesium clarification can be improved by adding magnesium salts such as $MgCl_2$ or dolomite limestone $CaMg(CO_3)_2$.[2] Dolomite lime $CaO.MgO$ is less effective because the MgO is resistant to slaking and does not dissolve at high pH to be precipitated as the gelatinous $Mg(OH)_2$.

B Flocculation

The rate of flocculation is proportional to the mean velocity gradient (G), the detention time, the concentration of particles and the particle size.

$$G = \left[\frac{C_D A v^3}{2 \nu V}\right]^{1/2}$$

where:

G = mean velocity gradient - sec^{-1}

C_D = drag coefficient

A = area of paddles - ft^2

v = mean velocity of the paddles relative to the water ft/sec

ν = kinematic viscosity of the water - ft^2/sec

V = volume of flocculator - ft^3

The most difficult task in calculating the velocity gradient is determining C_D and v. For flat blades made of wood C_D is taken to be 1.8. For metal blades the value of C_D is 1.2. The velocity of the paddles relative to the water is taken as $v_w = 0.75\, v_p$. Although this value will depend upon the geometric configuration of the flocculator and relative area of the blades to the flocculation basin. Without the use of stator blades the area of the paddles should not be greater than 15% to 20% of the cross-sectional area of the basin to prevent rolling of the water.[4] The flocculation process was characterized by Camp and Stein[5] in the following equation.

$$N = \frac{G}{6} N_1 N_2 (d_1 + d_2)^3$$

where:

N = number of collisions per unit volume per unit time.

G = mean velocity gradient - sec^{-1}

N_1 = number of particles originally in the water

d_1 = diameter of these particles

N_2 = number of floc particles in the water

d_2 = diameter of floc particles

Studies by Robeck[6] indicate N_1 is in the order of 10^6 particles per milliliter, and d_1 is in the range of $< 10\ \mu$. Floc particles on the other hand are often 100-200 μ. Therefore, d_1 can be omitted from the equation with little error.

C Sedimentation

The separation of dilute suspensions of flocculating particles is a function not only of the settling properties of the particles but also of the flocculating characteristics of the suspension. A fraction of the subsiding particles overtake and coalesce with smaller particles to form particles which settle at rates higher than the parent particles. The opportunity for contact increases with liquid depth. The removal efficiency is therefore dependent on both the clarification rate and depth. There is no satisfactory formulation available for evaluating the flocculation effect on sedimentation. This effect has to be evaluated by performing a settling-column analysis. Samples are withdrawn from several depths at different time intervals from the column. The solids concentration of each sample is determined and the fraction removed (relative to the initial concentration) is plotted on coordinates of time and depth to correspond to the sampling. Isoconcentration lines are then drawn to describe a depth-time ratio equal to the minimum average settling velocity of each fraction. Graphical summation of the removed fractions at the detention time and depth values compatible with the design clarification rate will give the overall removal efficiency to be expected. An illustrated example is presented in reference 3.

D Filtration[7]

1 Flow determination

The frictional headloss through a clear bed of media can be calculated by the following relationship which was developed experimentally by Rose.

$$h_L = f''' \cdot \frac{L}{\phi D_p} \cdot \frac{1}{e^4} \cdot \frac{Vs^2}{g}$$

where:

h_L = headloss, ft

f''' = bed depth, ft.

D_p = particle diameter, ft.

I = coefficient of nonshericity

e = porosity

Vs = average superficial velocity ft/sec

g = gravitational constant ft/sec^2

The dimensionless friction factor f''' is related to the drag coefficient C_d.

$$f''' = 1.067\ C_d$$

When flow is nonlaminar C_d is related to the Reynolds number as follows:

$$C_d = \frac{24}{N_{re}} + \frac{3}{N_{re}} + 0.34$$

For laminar flow, C_d approaches a value of $\frac{24}{N_{re}}$.

This expression applies strictly to flow through a clean bed. Once particulate matter has been deposited in the medium the relationship is not valid becuase particulate matter changes both the porosity of the bed and the shape of the bed particles. A valid expression to calculate the headloss across a filter in operation has not been developed. In practice, it is simply measured.

2 Removal mechanisms

The removal mechanisms described below have all been advanced as a possible mechanism of particle removal from a fluid during passage through a porous medium. Just as no exact mathematical model has been developed to describe filtration, no single mechanism can explain all the observed phenomena. It is a number of interacting mechanisms.

a Straining

The most logical mechanism for the removal of particulate matter from water is the deposition of particles too large to enter the openings of much smaller particles, which actually occurs. In addition, this mechanism implies that the size of particles removed is independent of the particle composition, which is not the case. Removal is dependent on the nature of the particle.

b **Sedimentation**

This theory considers each opening in the filter medium as a tiny sedimentation basin. If this were the case, the removal efficiency should be size dependent which is not actually observed. In addition, it offers no explanation why particles remain settled out, particularly since higher velocities will develop as the openings become constricted.

c **Inertial impaction**

The particulate matter has a greater density than the fluid transporting it. Consequently, as the streamlines of the fluid bend around an obstacle, the greater inertia of the denser particles forces them to the outside of the turning radius where they may inpinge on another obstacle. The removal efficiency of such a mechanism should increase with hydraulic loading which in practice does not occur.

d **Brownian movement**

Small colloidal particles could be forced out of the moving streams and into contact with a solid surface by random molecular motion but the effect is appreciable only for particles less than 2 μ in diameter.

e **Chance contact**

The chance contact of suspended particles with each other or with surfaces of the medium caused by flow through a constriction has been advanced as a logical mechanism, particularly if some other force preventing redispersal can be postulated. The probability of removal is thought to be directly proportional to the square of the particle diameter and inversely proportional to the cube of the diameter of the bed particles.

f **Diffusion**

Soluble and small suspended particles will migrate toward areas of low concentration which accounts for particulate removal into regions of the filter where flow is essentially zero. Colloidal particles will accumulate in regions of low shear even if they have to migrate against a concentration gradient. An equilibrium with the opposing shear forces probably develops. These forces are definitely operative but are they of sufficient magnitude to be of significance?

g **Van der Waals Forces**

These attractive forces vary inversely as the seventh power of their distance of separation and become significant only when the distance approaches the molecular diameter. Repulsion forces exist at these distances but are of a considerably smaller order of magnitude. Since cohesive forces are additive, the force is inversely proportional to the cube of the separation distance for large aggregates. Van der Waals Forces are not considered a dominant removal mechanism because of the rapid decrease of effect as distance increases. They may be significant, however, in preventing the redispersion of removed particulates.

h **Electrokinetic effects**

Dispersed particles in an aqueous environment with similar surface charges will exhibit a repulsive force. The presence of ions in the aqueous phase greatly influences these repulsive forces. Surface charges on the filter medium can determine the stability of the colloidal dispersion. A preferential removal of smaller negatively charged particles has been observed by investigators when a positive charge was imparted to the filter medium. The electrokinetic effect is opposite from what would be expected from a purely physical mechanism.

i Flocculation

The chance contact of suspended particles with each other and the filter medium provides an opportunity for floc formation and/or growth. These larger particles are mechanically held in the crevices of the filter until they are sheared off and driven deeper into the bed. It has been observed that filtration efficiency decreases with temperature due to the higher viscosity of the water and decreased flocculation.

j Biological activity

This is the predominant removal mechanism for slow sand filters. It is not significant in rapid sand filtration because of the velocity and frequency of backwash. Highly polluted waters with growth nutrients enhances biological activity and its influence on particulate removal.

k Chemical influences[8]

Recent work has shown that the chemical composition of the water had a significant effect on the removal of ferric floc. The phosphate ion was particularly influential. More work is required in this area.

The physical, chemical and biological influences on filtration have not been sufficiently defined; nor has a mathematical or empirical formula been developed to account for the changes occurring during filtration to permit the scientific design of a filter. Practical experience with successful installations and field investigation of new applications is still required to design an effective and economical filter plant.

III DESIGN CRITERIA

A Rapid Mix Compartment

1 Detention time 30-60 sec.

2 Design "G" 500 sec.^{-1}

3 Gt 30,000

B Flocculation Chamber

1 Detention time 10-30 min.

2 Design "G" 20-30 sec.^{-1}

3 Gt 54,000

Note: Hudson[9] advocates Gt values of up to 200,000 to produce a dense sludge.

C Sedimentation

1 Horizontal-flow basins

a Overflow rate - 700 GPD/FT^2

b Detention time 1-3 hrs.

c Used primarily for coagulants forming light hydroxide floc.

2 Up-flow clarifier

a Rise rate 0.5 - 1.5 GPM/FT^2

b Detention time 1-3 hrs.

c Sludge recirculation rate - up to 10% of design flow

d Used primarily for coagulants forming heavy precipitates capable of withstanding the high shear forces during recirculation.

D Filtration

1 Media

a Types

1) Sand

2) Anthracite coal

3) Garnet

- b Effective size
  - 1) Sand 0.35 - 0.55 mm
  - 2) Anthracite 0.75 - 1.80 mm
  - 3) Garnet 0.15 - 0.30 mm
- c Depth
  - 1) Sand - 6 in.
  - 2) Anthracite coal - 18 in.
  - 3) Garnet 1.5 - 3.0"
  - 4) Total 24 - 30 in.
- d Configuration
  - 1) Single
  - 2) Dual media
  - 3) Multi-media

2 Underdrains

- a Pipe lateral use 18 in. gravel to support media
- b Commercial underdrain systems, use minimum of 12 in. of gravel to support media.

3 Rate of filtration

- a Range 2 - 10 GPM/FT^2
- b Mean 3 GPM/FT^2

4 Backwash rate

- a Water 15 - 25 GPM/FT^2
- b Air - 4 cfm/FT^2

5 Mode of operation

- a Gravity
- b Pressure

IV OPERATING PARAMETERS

A Coagulation

1 Chemical addition sequence

- a pH or alkalinity adjustment
- b Primary coagulant
- c Coagulant aids
- d Filter aids

2 Point of chemical addition

- a Prior to rapid mix
- b During rapid mix
- c After rapid mix, prior to flocculation
- d After setting, prior to filtration

3 Optimum chemical dose determination

- a Empirical methods
  - 1) Jar tests
    - a) Floc size
    - b) Supernatant clarity
    - c) Appearance of first floc
    - d) Rate of settling floc
    - e) Filterability characteristics
    - f) Zeta potential
  - 2) Pilot filters to predict end of run conditions
  - 3) Production filters performance
    - a) Length of run
    - b) Product quality

b Analytical methods

1) Residual coagulant

2) Suspended solids

c Monitoring methods

1) Turbidity

a) Settled

b) Filtered

2) Conductivity

3) pH

B Filtration

1 Headloss termination criteria (usually 8 ft. of water for a gravity system)

2 Product quality criteria for termination (1.0 J.T.U.)

V PERFORMANCE IN A TERTIARY TREATMENT APPLICATION

A Alum Coagulation

1 Lebanon, Ohio

The Federal Water Pollution Control Administration has conducted a study of alum coagulation and filtration at Lebanon, Ohio. The final effluent from the activated sludge treatment plant had an average suspended solids concentration of 17 mg/l during the study period. The tertiary treatment consisted of chemical coagulation, flocculation, sedimentation in a horizontal-flow basin and gravity filtration. Table 1 indicates the range of variables investigated and the optimum value for each.

Under the optimum operating conditions it was possible to consistently produce a high quality water removing essentially all suspended solids and acid hydrolyzable phosphate. Filter runs have exceeded 30 hours to a terminal headloss of 8 feet at a filtration rate of 3 gpm/ft^2. Floc strengtheners were required to produce a filterable floc and prevent premature turbidity breakthrough with the dual media filters. Typical results achieved with optimum operating conditions are shown in Table 2.

2 Nassau County, New York

Pilot plant experience at Nassau County has also demonstrated the effectiveness of alum coagulation and filtration in a tertiary treatment application. Two desirable aspects of alum coagulation are the reduction in phosphate and pH, both are treatment objectives. Lime and iron salts were tried but found ineffective at Nassau, possibly because of the low alkalinity of the effluent. Alum dosages of 50-300 mg/l are used with a cationic polymer as a floc strengthener. A high-rate sludge blanket clarifier with sludge removal and a high-rate upflow clarifier with sludge recycle were found equally effective for clarification with a maximum rise rate of 0.5 gpm/ft^2 and detention time of 2 hours. Dual media filters consisting of 30" of 0.9 mm coal and 6" of 0.35 mm sand produced water with turbidities less than 0.4 J.T.U. when a small amount of alum (8-10 mg/l) and polymer (0.5 mg/l) were added after clarification and prior to filtration. The optimum filtration rate was found to be 4 gpm/ft^2 with filter runs of 8 to 24 hours.

B Lime Clarification

1 Lebanon, Ohio

Recently a 75 gpm single stage, up-flow sludge blanket clarifier was put on stream at the FWPCA Pilot Plant in Lebanon, Ohio. The lime addition of approximately 300 mg/l as $Ca(OH)_2$ is controlled by pH.

TABLE 1. OPTIMUM OPERATING CONDITIONS FOR TERTIARY CLARIFICATION USING ALUM COAGULATION FILTRATION AT LEBANON, OHIO

Variable	Range	Optimum Value
(1) Overflow Rate GPD/FT^2	300-900	700[1.]
(2) Alum Dose mg/l		
(a) Clarification	0-500	150
(b) Phosphate removal	0-500	300
(3) Type of Media		(c) Sand and coal dual media
(a) Sand		
(b) Coal		
(c) Sand and coal		
(4) Effective Size of Coal mm.	0.75-1.80	1.4 @ 4 GPM/FT^2
Sand Size 0.45 mm.		1.3 @ 3 GPM/FT^2
(5) Filtration Rate GPM/FT^2	2-5	4
(6) Floc Strengtheners		
(a) Activated silica	0-15 mg/l	3 mg/l
(b) Cationic polyelectrolyte C-7*	0-1.5	1.5 mg/l
(7) Source of AL^{+++} on Phosphate Removal		
(a) Alum	0-100%	No difference until pH exceeded 8.0. This occurred when 90% of AL^{+++} came from the aluminate
(b) Aluminate	0-100%	
(c) Both		

*A product of Rohm and Haas Company, Philadelphia, Pa.

[1] 700 GPD/FT^2 based on economic considerations; optimum solids removal efficiency occurred at the 300 GPD/FT^2 rate.

TABLE 2. TYPICAL RESULTS UNDER OPTIMUM OPERATING CONDITIONS

Process Stream	Suspended Solids, mg/l	Acid Hydrolyzable Phosphate, mg/l $PO_4^{\equiv}$	Turbidity J.T.U.
Secondary Effluent	45.6	22.4	12.2
Chemically Treated Settled Effluent	11.0(76%)*	2.2(90%)	1.5(88%)
Filtered Product	1.3(97%)	0.9(96%)	0.5(96%)

Operating Conditions:

Overflow Rate: 700 GPD/FT^2
Alum Dose: 82 mg/l
Aluminate Dose: 68 mg/l
Silica Dose: 3 mg/l

Filtration Rate: 3GPM/FT^2
Effective Size of Coal: 1.325 mm
Length of Run: 25 hours
Sludge Concentration: 1.2% by weight
Solids Concentration in Filter Backwash Water: 475 mg/l

* Denotes Removal Efficiency as %

The clarifier is operated at a rise rate of 1 gpm/ft^2. Dual media filters; consisting of 18 inches of 0.75 mm coal over 6 inches of 0.45 mm sand polish the clarifier effluent. The filters are operated at 2 gpm/ft^2 to a terminal headloss of 9 feet of water. Some preliminary operating results are shown in Table 3.

The suspended solids concentration of the settled sludge ranged from 1.5 to 2.5% solids by weight and was further concentrated to 12-15% solids by weight in a gravity thickener and additional drying on sand beds prior to disposal to a land fill.

C Jar Test Results

When effluent quality standards require the removal by phosphate in addition to clarification, significantly higher coagulant doses can be anticipated. A rough stoichiometry of 2 moles of AL^{+++} to 1 mole of P has been found empirically, to exist in tertiary treatment applications.

In addition, as shown in Figure 1 all the primary coagulants are equally effective in removing phosphate. These results were obtained from jar test studies[11] on a pilot plant activated sludge effluent.

VI INDUSTRIAL APPLICATIONS[12]

A Food Industry

1 Meat

Grit removal, flocculation and sedimentation will remove 50% of the total volatile solids and BOD and 75% of the suspended solids in a manure-bearing wastewater.. Alum coagulation removes an additional 2.5 mg/l of BOD and 1.5 mg/l of grease per mg/l of alum added. The chemical costs for 50% and 78.5% increased removal efficiency is \$0.64 to \$2.05 per 100 lb. BOD removal, respectively.

2 Poultry process wastes

Plain sedimentation can remove 17-28% of the BOD and 30-65% of the S.S.

TABLE 3. PRELIMINARY OPERATING RESULTS FOR LIME CLARIFICATION OF SECONDARY EFFLUENT

pH	TURBIDITY, J.T.U.			SUSPENDED SOLIDS mg/1		PHOSPHATE, PO_4^{-3} mg/1		LENGTH OF FILTER RUN (Hrs.)
	Influent	Settled	Filtered	Influent	Settled	Influent	Settled	
9.03	9.7	9.9	0.27	-	-	-	-	26
9.66	11.7	8.5	0.2	30.2	13.2	25.8	1.74	53.5
10.0	14.3	10.2	0.17	47.6	36.1	20.8	0.7	-

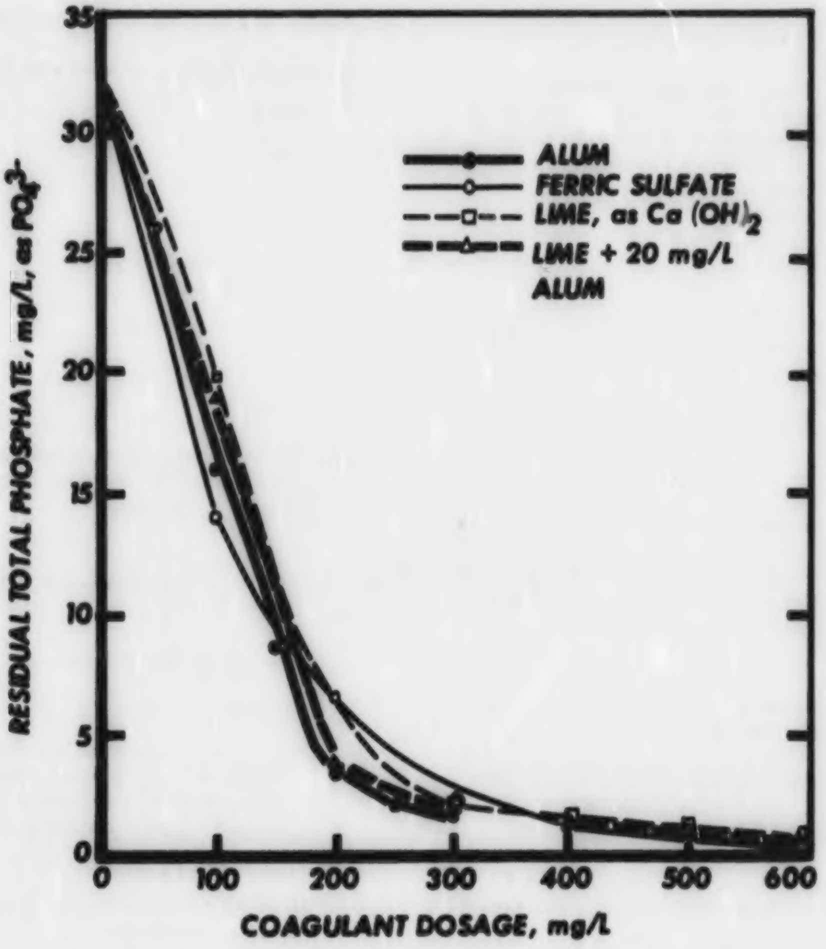

FIGURE 1 PHOSPHATE REMOVAL BY VARIOUS COAGULANTS

3 Canning wastes

Lime followed by either ferrous sulfate or alum can remove 50% of the BOD of pea and corn wastes and less than 20% of the BOD in fruit wastes. The treatment is considered a pretreatment to biological secondary treatment.

B Manufacturing Industries

1 Organic chemical wastes

Chemical coagulation and sedimentation is used to treat colloidal and oily wastes.

2 Pulp and paper industry

Without dispersing agents present, 70-80% of the solids in a paper making waste can be removed by plain sedimentation. Alum, activated silica and polyelectrolytes can increase the removal of suspended solids to better than 90%.

3 Textile industry

Chemical coagulation gives good results but is more expensive than biological treatment methods.

C Mineral Industries

1 Iron and steel industry

a Flume water clarification

Treatment consists of primary clarification in a scale pit followed by secondary clarification and oil removal. Lime, ferric sulfate and polyelectrolytes are used as coagulating agents. The water is usually suitable for reuse.

VII CAPITAL AND OPERATING COSTS OF CHEMICAL COAGULATION, SEDIMENTATION AND FILTRATION

A Capital Costs for a 10 MGD Chemical Coagulation, Sedimentation and Filtration Plant

Design Basis: Overflow Rate 700 gpd/ft^2
Filtration Rate 4 gpm/ft^2

1	Sedimentation basin: (including sludge collection system and appurtenances)	\$194,300
2	Influent pump:	5,400
3	Chemical feed equipment	48,000
4	Land at \$10,000/acre	15,000
5	Filters:	462,500
6	Total capital costs	725,800

B Operating and Maintenance Costs

1	Amortization, 20 years at 4%:	53,500
2	Labor:	26,000
3	Chemicals:	172,300
4	Power:	9,125
5	Maintenance:	3,630
6	General overhead and Administration:	5,820
7	Total annual cost:	270,375

C Unit Cost for Clarification and Phosphate Removal

Based on annual water production of 3.65×10^9 gallons the unit cost of water is $0.074/1,000 gal.

VIII ULTIMATE DISPOSAL OF CHEMICAL SLUDGES

A General Disposal Techniques

1 Lagooning

One of the most widely practiced methods of disposal. It is to be considered a stopgap method because it creates both a nuisance and a liability.

2 Sludge drying beds

Must be evaluated in terms of local climatic conditions such as evaporation rate and mean annual rainfall. As much as 3-5 acres per mgd of capacity may be required.

3 Disposal to a waste sump

This is often a more "ultimate" type of disposal depending upon future land use and degree of isolation from the surrounding environment.

a Costs

Mode of Transportation	Cost in Mills Per Ton Mile
Rail	34-40
Truck	15
Barge	4
Pipe	2

Pumping a distance of 250 miles is considered reasonable and a minimum scour velocity of 5 ft/sec is recommended.

Power costs are 12% of the total cost per year if the optimum hydraulic pipe size is installed.

4 Vacuum filtration

Sludges from coagulation facilities are often thin and generally difficult to dewater. Intermediate dewatering operations such as thickening should be considered to reduce the volumetric load on the filter. Significant savings can be achieved by altering the process conditions to yield a sludge of superior dewatering characteristics.

Vacuum filtration is attractive because of small space requirements, all year operation, widespread applicability, and satisfactory properties of the cake.

5 Recovery and reuse

Recovery and reuse of chemical sludges offers the most promising solution to the disposal problem because significant savings in chemical costs can be realized and the ultimate disposal problem is minimized.

B Recovery and Reuse of Aluminum Hydroxide Sludge

Since aluminum hydroxide is amphoteric it will ionize in both acid and base. The acid recovery scheme is not practical for wastewater treatment because of the difficulty associated with separating the phosphates from the recovered aluminum. The alkaline method of alum recovery is the most promising solution and merits further investigation. The recovery scheme involves dissolving the aluminum hydroxide and adsorbed phosphates by raising the pH to 11.9 with NaOH. The reaction converts the aluminum hydroxide to sodium aluminate. Calcium chloride is then added which reacts with the phosphates to form the insoluble precipitate of tricalcium phosphate. The calcium phosphate is separated by sedimentation and/or filtration and the comparatively phosphate-free sodium aluminate solution is reused. The reactions for this process are as follows:

$$\mathrm{Al(OH)_3 \cdot [PO_4^{=}] + 4\,NaOH \xrightarrow{pH = 11.9} NaAlO_2 + Na_3PO_4 + 2H_2O + 3OH^-}$$

$$\mathrm{NaAlO_2 + Na_3PO_4 + 3CaCl_2 \longrightarrow NaAlO_2 + Ca_3(PO_4)_2 + 6NaCl}$$

Culp, et al,(13) in laboratory tests achieved 75% recovery of the feed alum by sedimentation and 85% by filtration compared to the 93% recovery reported by Lea, et al(14). Lime can also be used to recover alum. The lime not only raises the pH but also supplies the necessary calcium to precipitate the phosphates. Although the percent recovery is less with lime (35%); the cost of chemicals to regenerate a ton of alum is lower. If alum recovery proves successful, treatment costs can be reduced to $0.05/1,000 gal. The problem with alum recovery to date has been the decreased efficiency of the recycled sodium aluminate in removing phosphate.

C Recovery and Reuse of Lime Sludges

A two stage lime precipitation treatment system including facilities to recover and reuse the lime is being installed at the FWPCA Pilot Plant in Washington, D.C. Lime is added with or without a magnesium salt or other coagulant aid as required to the first stage precipitator. At a pH of 11.3 the calcium and magnesium phosphates, sulfates and carbonates are precipitated together with the organic sludge. Compressed stack gases which are rich in carbon dioxide are usually used to recarbonate the waste stream as it enters the second stage clarifier. The excess calcium is precipitated as almost pure calcium carbonate to a pH of 9.0. Sludges from the two stages are dewatered by centrifugation or vacuum filtration and then calcined in a multiple hearth furnace at 1800°F to produce calcium oxide. Calcining destroys the organic matter, but phosphates accumulate as inert tricalcium phosphate along with other inert minerals which must be wasted. The calcium oxide is then slaked and reused. The chemical cost savings realized by recalcining versus purchase of fresh lime is negligible. The significant cost savings are the reduced disposal costs. Cost estimates for the Washington, D.C. Plant at the 250 mgd size indicate that savings of up to $0.02/1,000 gal. can be realized.

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ice on the pond. The lowest temperature of the pond effluent was 48°F, and this wa recorded after 5 days during which temperatures were below 0°F. Facultative operation of the lagoon produced small amounts of algae in the pond during the summer period, but no odor problems were noted during the operation of this lagoon.

Operational data from the facultative effluent polishing lagoon in Decatur, Ill., also shown in Table 7-5, indicate that BOD and suspended solids removals averaged 40 percent and 49 percent, respectively, while operating under an organic surface loading of 218 lbs BOD/acre/day (6). As seen in Table 7-5, the facultative lagoon at Springfield, Mo., receives much higher surface organic loadings (approximately 1,290 lbs. BOD/acre/day) and still performs creditably, with average BOD and suspended solids removals of 64 and 62 percent, respectively (7).

A major disadvantage of using a facultative lagoon is the fact that the effluent will have a minimal dissolved oxygen content. Springfield, Mo., solved this problem by using cascade aeration (See Chapter 8, Section 8.2 on Post-Aeration). The effluent from the polishing lagoon flows over a series of 5 weirs with a total drop of 75 inches. The average dissolved oxygen in the effluent (September, 1970 through March, 1971) was 7.0 mg/l with a minimum and maximum, respectively, of 4.0 and 9.9 mg/l (7).

7.3 Microstraining

Microstraining has application in effluent polishing chiefly as a method of removing additional suspended solids (and their associated BOD) from wastewater treatment plant effluents. A typical unit is shown in Figure 7-2. (8). The microstrainer consists of a rotating drum with a peripheral screen. Influent wastewater enters the drum internally and passes radially outward through the screen, with deposition of solids on the inner surface of the drum screen. At the top of the drum pressure jets remove the deposited solids. This backwash water is then collected and returned to the head of the plant.

The screens employed in microstrainers have extremely small openings and are made from a variety of metals and plastics. Individual manufacturers have specific designs and sizes for the particular needs of any potential installation. One manufacturer offers the following grades of microfabric (9):

Table 7-6

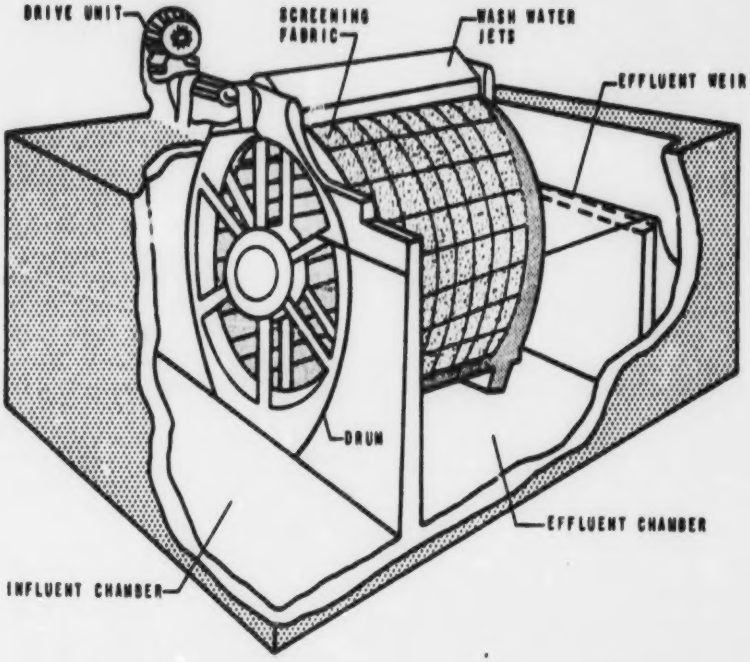
Microstrainer Fabric Sizes

Opening microns	No. of Openings per sq.in.
23	165,000
35	80,000
60	60,000

This copy taken from the "Process Design Manual for Upgrading Existing Wastewater Treatment Plants" prepared by Roy F. Weston, Inc. under Contract 14-12-933 with the EPA Technology Transfer Program.

FIGURE 7-2

TYPICAL MICROSTRAINER UNIT (8)


DRIVE UNIT
SCREENING FABRIC
WASH WATER JETS
EFFLUENT WEIR
DRUM
EFFLUENT CHAMBER
INFLUENT CHAMBER

The weave and shape of individual fabric wires are such that they allow the water from the backwashing jets to penetrate the screen and remove the solids mat which forms on the inside of the screen during its passage through the feed stream. Bodien and Stenburg (8) have noted that only about one-half of the applied washwater actually penetrates the screen; the rest flows down the outer perimeter into the effluent chamber. Previously-strained effluent can be used as washwater.

Although the microstrainers have small openings, the openings themselves cannot account for the removal efficiency of the unit. Actually, the mat of previously trapped solids provides the fine filtration which characterizes the unit performance. This being the case, Lynam et al (10) showed quantitatively that the slower the rate of drum rotation, the better the product water. Another factor which becomes important in light of this mat phenomenon is the nature of the solids applied to the microstraining process. For example Lynam et al were unsuccessful in filtering the resulting chemical floc when secondary effluent was coagulated ahead of the microstrainer unit.

As a section of the screen passes through its cycle, it becomes clogged rapidly as the solids mat forms. The continuous cleansing afforded by the backwashing jets at the apex of its travel must be augmented in some way to prevent the buildup of screen-clogging slimes over a period of time. Ultraviolet light placed in close proximity to the screen has been somewhat successful in slowing the development of these slimes. In general, however, units must be taken out of service on a regular basis (once/week, for example) to have the metal screens cleaned with a chlorine solution. In some instances, a similar cleaning with an acid solution may be required on occasion to deal with iron or manganese buildup on the metal screen. In cases where oil and grease problems occur, a hot water and/or steam treatment can be used to remove these materials from the screen.

One of the advantages of using a microstrainer is its low head requirement. It is, therefore, advantageous to transfer secondary effluent, without pumping, to a tertiary microstraining installation in order to minimize the shear forced imparted to the fragile biological floc. Head loss through the microstraining unit, including inlet and outlet structures is about 12 to 18 inches (9). Across the screen, a 6-inch limit is usually imposed at peak flows. Head losses in excess of this value are prevented by bypass weirs. Head loss buildup is reduced by increasing the rate of drum rotation and by increasing the pressure and flow of the backwashing jets. These adjustments can be made manually or automatically.

Other operating parameters include the hydraulic and solids loading on the unit. Lynam et al (10) found that the solids loading was the limiting factor in microstraining of activated sludge secondary effluent. Maximum capacity was found to be 0.88 lbs./day/sq.ft. at a hydraulic loading of 6.6 gpm/sq.ft. Excessively high solids during upset periods can reduce the capacity of the unit drastically from design levels. Chlorination immediately ahead of microscreening units should be avoided to protect the screens.

Operational data from various installations are presented in Table 7-7. The microstrainers using 23-micron fabric exhibited average solids removals ranging from 57 to 89 percent,

441-563 O - 71 - 10

Table 7-7

Microstrainer Operational Data

Location	Plant Size mgd	Feed Type	Fabric Opening microns	SS Removal percent	Effluent SS mg/l	BOD Removal percent	Effluent BOD mg/l	Backwash % of flow	Reference
Brampton, Ontario	0.1	A.S.[1] Effluent	23	57	—	54	—	—	9
Lebanon, Ohio	Pilot	A.S. Effluent	23	89	1.9	81	—	5.3	8
	Pilot	A.S. Effluent	35	73	7.3	61	—	5.0	8
Chicago, Illinois	3.0	A.S. Effluent	23	71	≈3.0	74	≈3.0	3.0	10
Luton, England	3.6	Effluent from A.S. and T.F.[2]	35	55	7.3	30	—	3.0	11
Bracknell, England	7.2	T.F. Effluent	35	66	5.7	32	8.4	—	9,11

[1] A.S. - Activated Sludge
[2] T.F. - Trickling Filter

while the 35-micron fabric exhibited removals of 55 to 73 percent. In practice, the coarser 35-micron fabric is generally used for the removal of coarse solids. Maintenance of microstrainers can be quite costly, since most units will require cleaning at least once a week as previously mentioned.

Typical design parameters for microstrainers are presented in Table 7-8. These parameters must be evaluated to determine the proper microstrainer design for existing conditions.

Table 7-8

Typical Microstrainer Design Parameters

Parameter	Value
Drum Speed, rpm	0.7 - 4.3
Filter Fabric, microns	23 and 35
Average Hydraulic Loading[1], gal/sq.ft./hr.	
23-micron fabric	600
35-micron fabric	800
Backwash Pressure, psi	20 - 80
Amount of Backwash Water, percent of average flow	3 - 6
Maximum Hydraulic Loss through Screen, inches	6

[1] Based on submerged screen area.

Sources: Diaper (9) and Lynam (10).

7.4 Multi-media, Coarse-media, and Moving-bed Filters

Historically, sand filtration has not been an efficient method of polishing secondary treatment plant effluent because of low application rates, high head losses, and the need for frequent backwashing. This is largely because the normal backwashing of a sand filter results in a size-graded filter with the finest grains in the upper layers. The resulting stratification removes the bulk of the suspended matter in the upper levels, with a consequent inefficient use of the remaining depth of the filter.

However, developments in mixed, multi-media, and deep-bed coarse-media filters have necessitated a re-evaluation of the role of filtration in effluent polishing. In general, these modifications permit deeper penetration of the media by the suspended and colloidal contaminants; thus, there is a more effective utilization of the filter depth as compared to conventional sand filters. The increased utilization of filter depth is somewhat offset by the fact that increased backwashing rates and larger quantities of washwater are required to backwash the media properly.

Deep-bed coarse-media sand filters are a modification of the typical rapid sand filter. The deep-bed filter has a minimum of 4 feet of media as compared to the 2.5 feet or less

PROCESS INSTRUMENTATION

I INTRODUCTION

The purpose of this presentation is to acquaint the trainee with some of the instruments that are available today for possible on-line applications to conventional sewage treatment processes and to describe briefly how they work. Finally, a few instruments which are in advanced stages of development will be touched on.

II ON-LINE INSTRUMENTS

Figure 1 shows where some of these instruments have been or could be applied to a typical activated sludge plant. There are many other plant configurations which could be used and there are many other possible instrument applications. This diagram is for illustrative purposes only. It is not intended to show the best or recommended way of doing any of these things. Where pieces of equipment are identified by brand name, manufacturer, or sketch, no endorsement is implied. Usually, similar instruments are available from several manufacturers.

In the detailed descriptions which follow, these instruments are in the same numerical order that is shown on this schematic diagram. Where only a signal line is shown running from a sensor to an instrument, the output is only a meter reading or a chart record. Where a control signal line is shown running from the instrument to some other component, such as a valve or motor, a process control function is indicated.

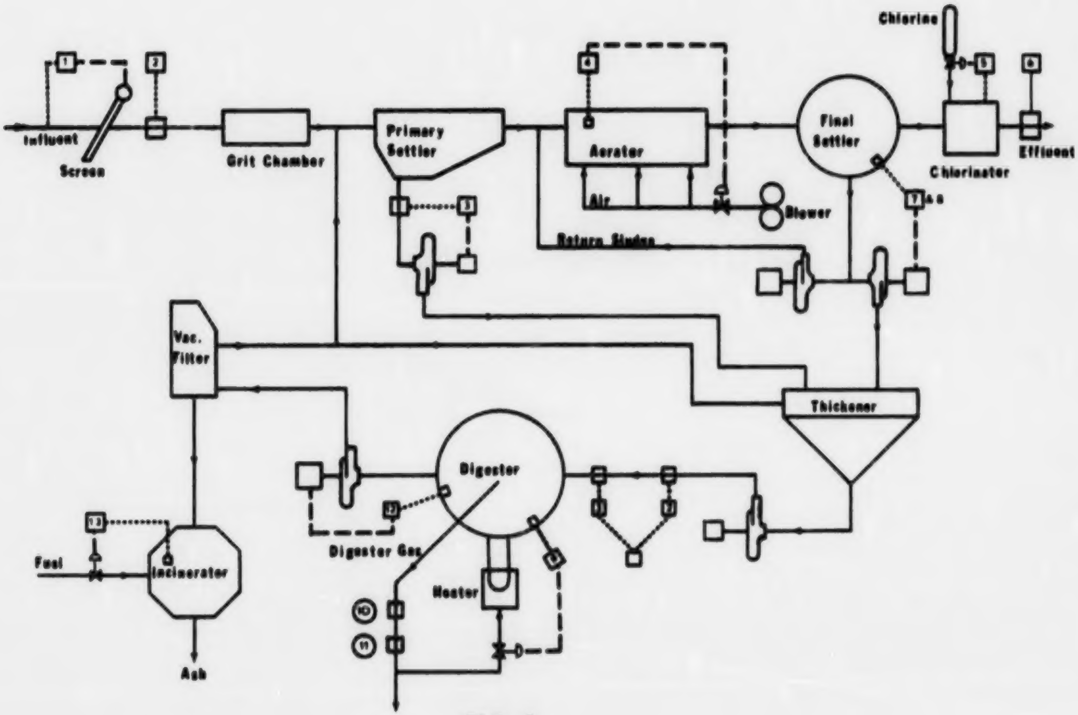

Chlorine
1
2
4
5
6
Influent
Screen
Grit Chamber
Primary Settler
Aerator
Final Settler
Chlorinator
Effluent
3
Air
Blower
7 & 8
Return Sludge
Vac. Filter
Thickener
Digester
12
13
Fuel
Incinerator
Digester Gas
Heater
10
11
Ash

FIG. 1

III INDIVIDUAL INSTRUMENTS AND CONTROLS

A Bar Screen Control (Figure 2)

This device measures the differential level upstream and downstream of the bar screen by bubbling air through tubes immersed in the flowing sewage and measuring the back pressure created by these levels. This differential can be read out or recorded directly in inches of water loss across the bar screen. The cleaning mechanism on the screen can be automatically actuated when this loss exceeds some present value. Additional electrical contacts can be provided to actuate alarms when unusually high head losses occur.

Such a device will cost from $500 to $600. One manufacturer is the Foxboro Company.

BAR SCREEN CONTROL

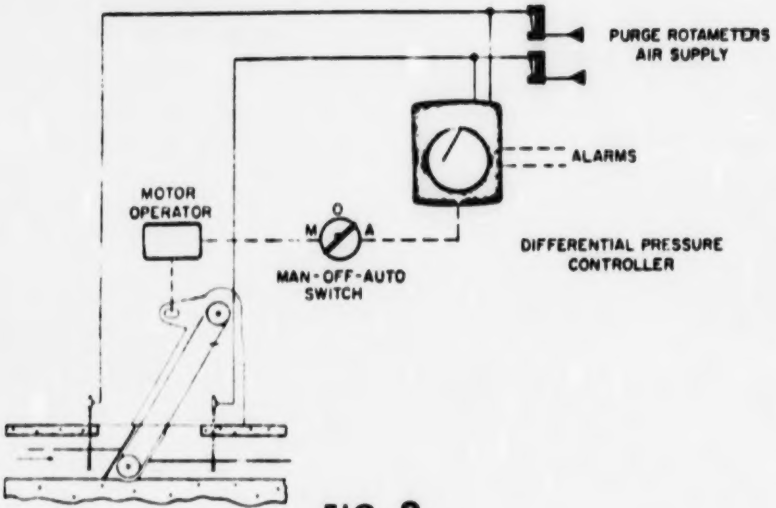

PURGE ROTAMETERS
AIR SUPPLY
ALARMS
MOTOR OPERATOR
O
M
A
MAN-OFF-AUTO SWITCH
DIFFERENTIAL PRESSURE CONTROLLER

FIG. 2

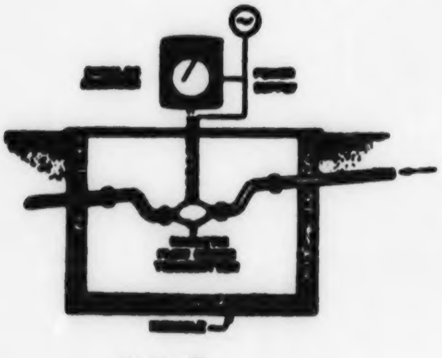


FIG.3

B Raw Sewage Flow (Figure 3)

There are a number of ways to measure liquid flow, including flumes, venturi meters, orifice plates and magnetic flow meters. This figure illustrates the magnetic flow meter which is well suited to this application since it does not require any pressure taps and is unaffected by entrained solids or trash. It will work on pressure or gravity lines; the only requirement being that the line is completely filled with sewage at the point of measurement.

The meter works on the principle of electro magnetic induction which says that a conductor moving at right angles to a magnetic field generates a voltage directly proportional to the velocity of the conductor. The meter is made up of a transmitter and a receiver (recorder or indicator). The transmitter consists of a length of nonmagnetic pipe, a section of which is enclosed by electromagnet. Two metallic electrodes, directly opposite each other and flush with the inside of the pipe, detect the low-level voltage signal generated by the moving sewage. This signal is transmitted to the receiver where it is read or recorded on a uniform scale. Minimum full-scale velocity is about 2.5 feet per second. Maximum velocity is usually limited by hydraulic considerations. Such meters are sold by Fischer and Porter, and by Foxboro. Prices depend upon pipe size, with a 6" meter and recorder costing about $3,000.

C Primary Sludge Density Meter (Figure 4)

The benefits derived from pumping thick sludge are obvious and several methods are used to control the density of pumped primary sludge. In small plants the sludge is often visually observed, while in larger plants programmed timers are used. Another method is the use of a sludge density meter which will provide continuous density analysis and pumping control.

These meters consist of a Source Head, a Detector Unit, and an Amplifier. The source head and the detector are mounted opposite each other on the sludge piping. The source head contains a quantity of some radioisotope such as Cesium-137. Gamma rays from this source pass through the pipe, through the sludge where they are absorbed in proportion to the density or percent solids, and into the detector unit. The rays reaching the detector produce a current which is inversely proportional to the sludge density. This current is amplified in the amplifier and passed on to a measuring instrument and controller where the results can be read directly in density units.

At present these meters are recommended only for primary sludge. Prices vary with

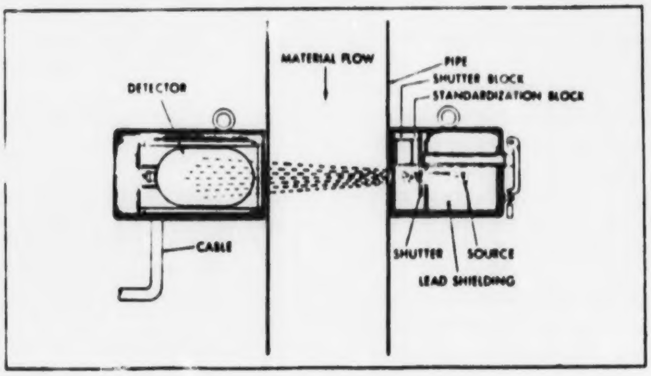
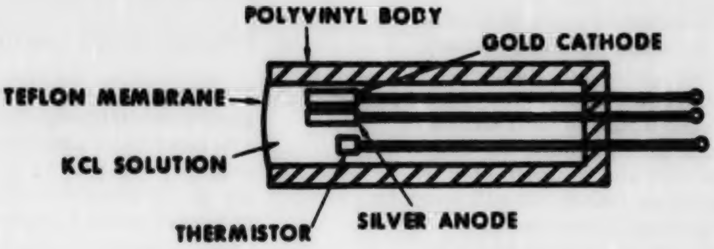
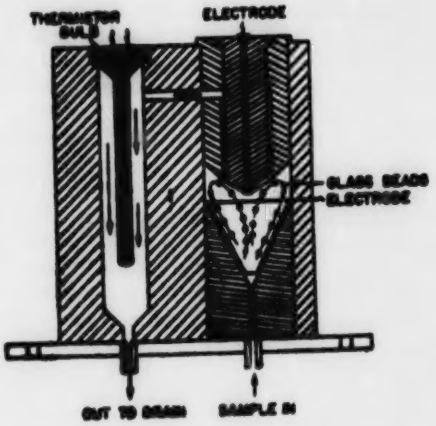


FIG.4



DISSOLVED OXYGEN SENSOR

FIG. 5



CELL DETAILS

FIG. 6

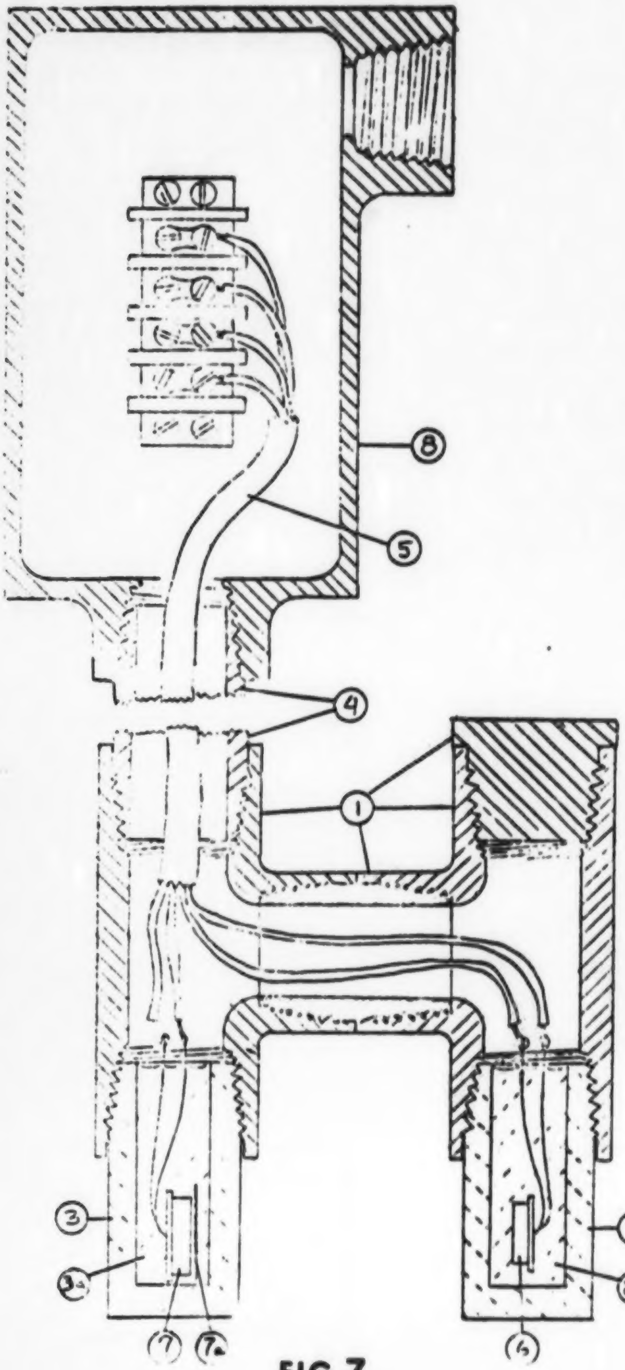
pipe size, but will run about $4,000 for a 6' line. Three manufacturers are: Industrial Nucleonics, Ohmart and Nuclear-Chicago.

D Dissolved Oxygen Meter (Figure 5)

One of the quantities which is of interest in the activated sludge process is the level of dissolved oxygen in the mixed liquor in the aerator. Since the cost of aeration is the largest single cost in the activated sludge process, automatic control of the DO level could result in a significant cost reduction.

There are several different brands of DO meters available which use a polarographic electrode for a sensor. Oxygen is reduced at the cathode surface causing a current to flow which is proportional to the partial pressure of oxygen in the liquid being tested. This current signal is then fed into an amplifier, the output of which can be used to drive recording or control devices.

Several different pairs of metals are used for the sensor depending on the manufacturer. Two of these are, silver anode and gold cathode, lead anode and platinum cathode. In some instruments the electrodes are separated from the sample by a thin Teflon membrane, which is permeable to gases. When this construction is used, a supporting electrolyte such as potassium chloride is used. These instruments are available for portable use and fixed installations. They are sold by Bechman, Fairchild, Robertshaw, Union Carbide, and Weston and Stack. Those for fixed installation cost about $1,400.



1. Body
2. Light Source Holder
2a. Potting Compound
3. Photocell Holder
3a. Potting Compound
4. Connecting/Mounting Shaft
5. Connecting Cable
6. Light Source
7. Photocell
7a. Photocell Filter
8. Junction Box

Sludge Level Controller
Probe Cross Section

FIG.7

E Chlorine Residual (Figure 6)

This is a schematic drawing of a Free Residual Chlorine Analyzer Cell made by Foxboro. The center electrode is platinum and the outer electrode is copper. The glass beads are used to keep the electrodes clean. The sample stream enters the electrode chamber where a voltage is generated by the fluid across the electrodes. The sample then passes to the adjacent chamber where it contacts a thermistor which provides temperature compensation. The electrical signal can then be used to operate a recorder or control a chlorinator. For satisfactory operation, the pH should be between 6 and 8.

This instrument, with a controller, costs about $2,300.

F Turbidity Meter

The clarity of the effluent can be measured as turbidity. Three manufacturers that make turbidity measuring instruments are Beckman, Foxboro and Union Carbide. All three of these use a light source and measure either transmitted light, scattered light or reflected light. For example, the Union Carbide meter uses a lamp, the beam of which is split at right angles and directed toward two photo conductive cells. The path to one cell is about 1-1/2 times as long as the other. The longest path is a little less than 1". This unequal path length is used to compensate for dirt on the detector windows. As one detector sees a decrease in light due to either dirt buildup or turbidity, it causes an increase in lamp voltage to bring the signal back to its initial strength. The difference between the signals from the two detectors is then read as turbidity. Prices for these meters range from about $1,400 to $2,000.

G Sludge Blanket

Two instruments which are quite different, but yet provide the same information, are shown in Figures 7 and 8. Neither one will indicate what the sludge level is at any given time, but both are intended to detect the sludge level when it reaches some predetermined point.

The first (Figure 7) is the American Bowser Corporation Sludge Level Probe. It is intended for use with a timing device and sensitivity selector to control sludge pumping. The components of the probe are identified in the illustration. One feature, which is not shown, is that the light source is a light emitting diode which is immune to shock and permits the use of a filter to restrict the response to sunlight. A timer is used in conjunction with the probe to prevent frequent on-off operations of the pump. This probe, together with the necessary power supply, relays, timer, etc., costs about $1,200.

The second sludge blanket level instrument is shown in Figure 8. This is the Sensall Interface Sensor, manufactured by the National Sonics Corporation. The sensor consists of two piezoelectric crystals separated by the transmitter crystal, directed through the liquid and picked up by the receiver crystal and then amplified by the control unit.

The sensor may be used either in a reflection mode or attenuation mode. To operate in a reflection mode, it must be mounted about 10° to the horizontal as shown here. When the ultrasonic signal passes the interface, most of the signal is reflected and refracted so that only a small amount reaches the receiver. If the interface is not between the two crystals, the greater part of the signal will reach the receiver.

Usually liquids with entrained solids have more sonic attenuation than clear liquids. If the attenuation of the sludge is sufficiently greater than that of the liquid above it, the attenuation mode can be used to determine

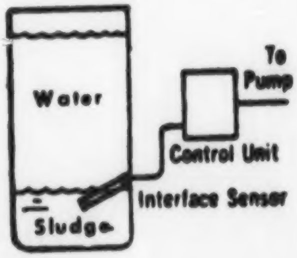
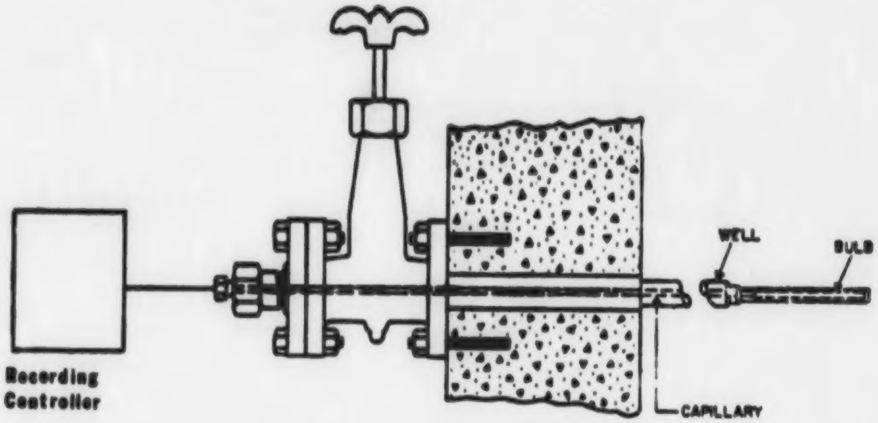


FIG.8



Thermometer Well and Bulb

FIG. 9

the type of liquid present at the sensor. However, this mode of operation is not recommended for sewage sludge.

This sensor together with the control unit sells for about $350.

H Mass Flow of Sludge

Measuring and recording of mass flow of sludge to the digester, may be accomplished by electronically combining the output signals from a magnetic flow meter and a sludge density meter and feeding the resultant signal into a recorder. The magnetic flow meter and sludge density meters would be essentially the same as those described above for measuring raw sewage flow and primary sludge density. The manufacturers of either the flow meters or density meters will assemble the meters, the recorder, and the necessary electronics for about $10, 000.

I Digester Temperature (Figure 9)

Probably the most familiar application of automatic control is that of digester temperature. These instruments are available from a number of manufacturers. This one shown here consists of a sensing element in a protective well with a capillary tube leading to the recording controller. The signal from the controller to the fuel control valve can be either pneumatic or electric. Often the temperature controller will be integrated with the burner and will be separated from the recording function.

An instrument for local recording and control of digester temperature will cost around $700.

J Digester Gas Flow (Figure 10)

Since the quantity of gas produced by a sludge digester is a good index of digester efficiency, it may be desirable to measure and record gas production rate. This can be done with several different types of meters, including a simple orifice plate meter as shown here. The pressure differential across the orifice plate is transmitted for remote reading. Such meters are available from Foxboro, Meriam, and others. The cost is about $400.

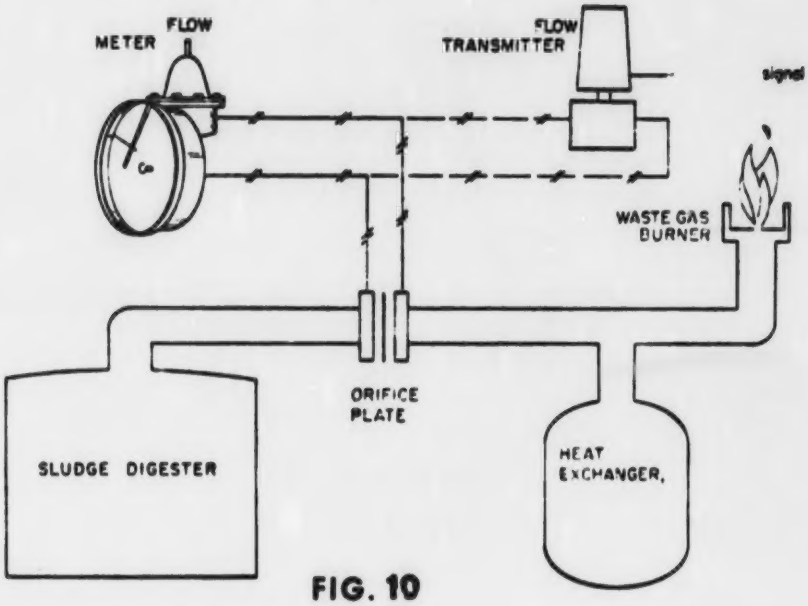
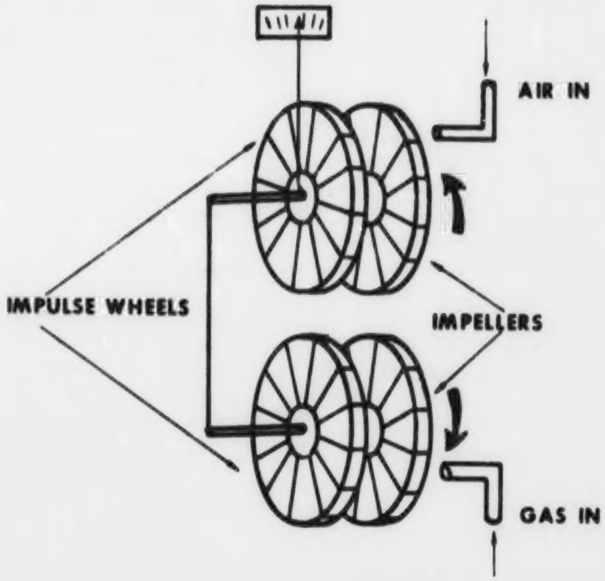

METER
FLOW
FLOW
TRANSMITTER
signal
WASTE GAS
BURNER
ORIFICE
PLATE
SLUDGE DIGESTER
HEAT
EXCHANGER,

FIG. 10


AIR IN
IMPULSE WHEELS
IMPELLERS
GAS IN

RANAREX INDICATOR

FIG. 11

K Digester Gas Analysis (Figure 11)

Typically, digester gas is about 65% methane and 35% carbon dioxide. Problems with digestion may be signaled by a sudden rise in CO_2 before total gas production falls off. Therefore, it may be desirable to monitor the CO_2 content of digester gas.

There are a number of ways to measure CO_2, among which is the unique instrument shown here. It is called Ranarex and is made by the Ranarex Department of the Permutit Company. It is a mechanical device which actually measures the specific gravity of the gas. It will, therefore, provide analysis of a two-phase gas system. The figure shows the operating principle. The instrument consists of two hollow cylindrical chambers each of which contains a motor-driven impeller and an impulse wheel. The lower impeller draws in a continuous sample of gas and discharges it against the vanes of the companion impulse wheel creating a torque proportional to the density of the gas.

Similarly the upper impeller draws in a continuous sample of ambient air and discharges it at the same speed, but in opposite direction to the gas creating an opposing torque proportional to air density. The difference between the opposing torques is a measure of the specific gravity and is transmitted through impulse wheel pivot shafts to a pointer and a recording chart pen.

There are a few conditions which must be met. The gas should be essentially at atmospheric pressure and ambient temperature and both gas and air should have the same humidity. This last condition is easily met by saturating the air in a built-in humidifier. Some advantages claimed for this instrument are rapid response, rugged construction, simplicity, and low installation cost. The price including recorder is about $1,200.

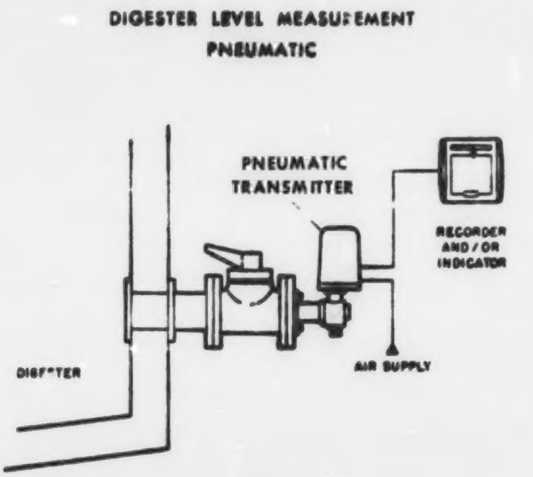


FIG.12

L Digester Level (Figure 12)

Accurate knowledge of liquid level is essential in digester operation. The instrument shown here will permit remote reading or recording of that level. This particular instrument operates by balancing an air pressure on one side of a diaphragm against the pressure of the digester liquid on the other side. By pneumatic transmission the level signal may be used to indicate or record digester level or to sound alarms. To facilitate service, a valve is installed between the digester and the instrument. These instruments are available from Foxboro and other manufacturers. With a recorder, they cost about $600.

M Incinerator Temperature

Incinerator temperature must be controlled to insure destruction of the sludge without exceeding safe upper limits with resultant damage to brickwork and other incinerator parts. Such controls are usually integrated with the burners and are often supplied with incinerators. Two manufacturers of temperature controlled burners are Eclipse Fuel Engineering and North American Manufacturing.

IV NEW INSTRUMENTS

There are several instruments of interest, which are not shown on the flow diagram (Figure 1). Some have only recently been put into production.

Figure 13 shows the Arthur Respirometer now called the Biomand I by the manufacturer, Phipps and Bird, Inc. This instrument will measure oxygen demand in wastewater. The schematic shows the principle of operation.

The sample is introduced manually and the automatic cycle started. When the cycle begins, the vent solenoid (6) opens, venting the system, allowing the manometer fluid to equalize and the recorder pen to return to the base line. The drain solenoid (1) opens, allowing the sample chamber to drain. The pumps (2) (12) stop at the end of the drain cycle. The drain solenoid closes and the fill solenoid (3) opens, allowing the sample to enter the sample chamber. The fill solenoid closes, the pumps restart, and the system vent closes.

As the available oxygen is used, the pressure drop in the closed system (7) allows the manometer fluid to flow from the open side. As the float (8) which is attached to a transducer (9) drops, a signal is sent to the recorder (10) which reads directly in milliliters of oxygen demand. It is claimed that a procedure which takes about four hours gives results which can be correlated with a 5-day BOD for a particular wastewater. This instrument costs about $7, 500.

Figure 14 is a schematic diagram of the "Continuous TOC Monitor" made by Automated Environmental Systems, Inc. It is intended for continuous on-line operation and comes equipped with either an on-line blender or on-line filter. The method of operation can be seen from this diagram. A constant head tank provides a continuous sample of the water to be analyzed. From this tank the water enters the top of the nitrogen sparge column where it is treated with hydrochloric acid to maintain a pH of about 3. The sparge column, essentially a counter-current gas stripper, removes CO_2

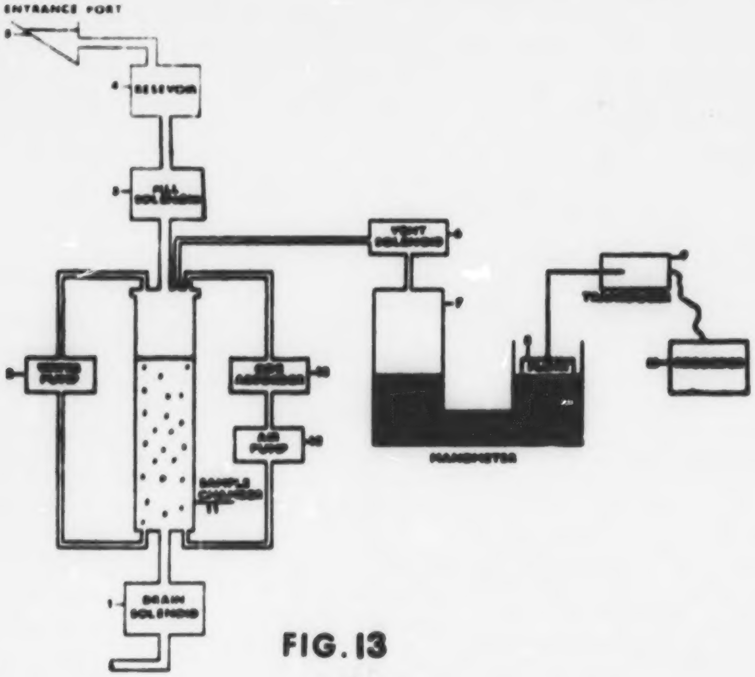
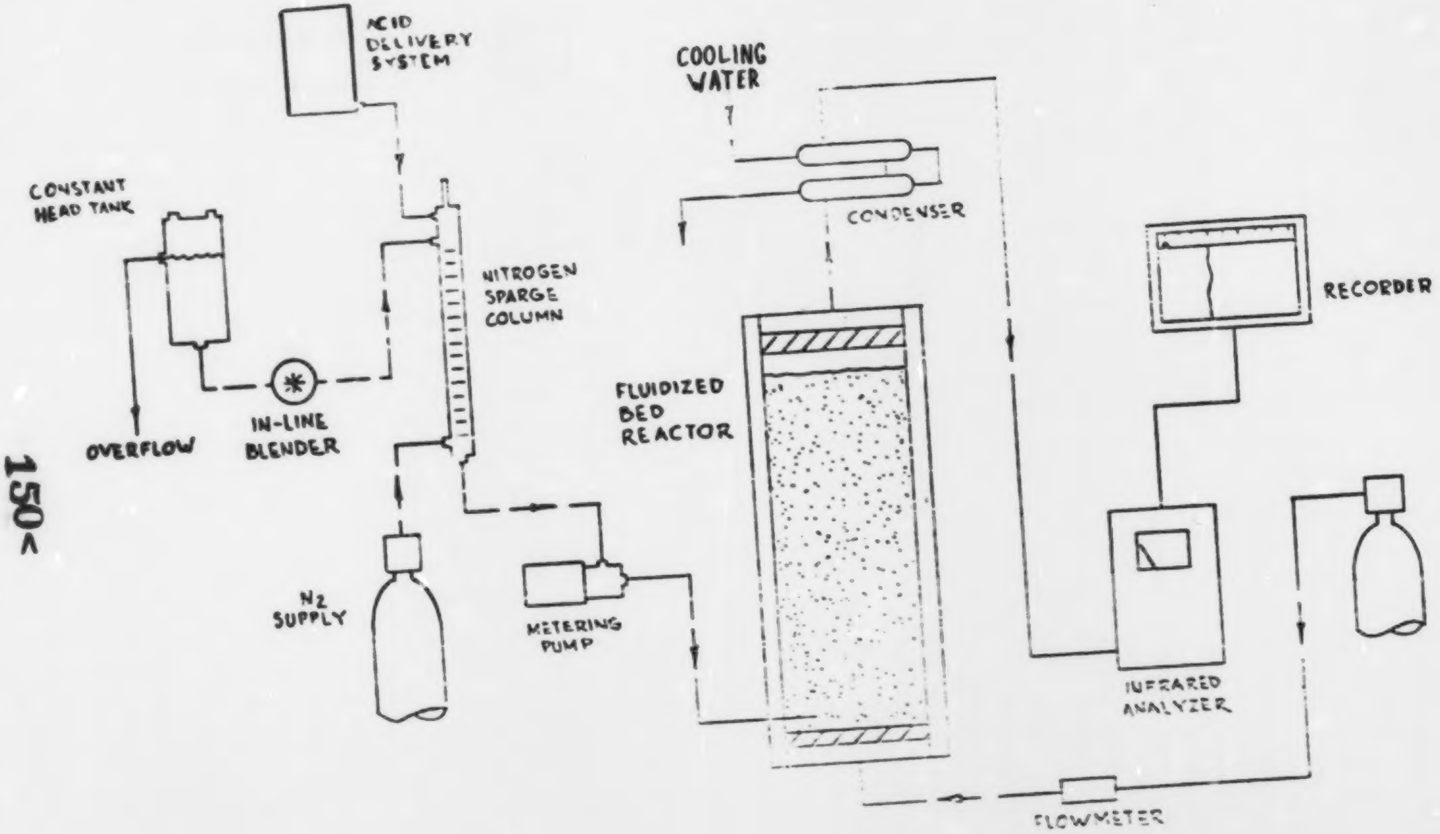


FIG. 13


ACID DELIVERY SYSTEM
COOLING WATER
CONSTANT HEAD TANK
NITROGEN SPARGE COLUMN
CONDENSER
RECORDER
FLUIDIZED BED REACTOR
OVERFLOW
IN-LINE BLENDER
N2 SUPPLY
METERING PUMP
INFRARED ANALYZER
FLOWMETER

CONTINUOUS TOC MONITOR — FLOW DIAGRAM
FIG.14

from the acidified sample, yielding an effluent free of inorganic carbon.

The carbon-free sample is then metered into a fluidized bed reactor by a metering pump. The bed temperature is 850°C. Oxygen is supplied at ten times the stoichiometric requirement to oxidize the maximum amount of TOC which may be present. The oxygen and the vaporized sample provide the gas for fluidization of the bed. In the reactor, organic carbon in the sample is oxidized to carbon dioxide. The reactor's effluent is a gas stream of water, CO_2 and O_2, plus oxidation products of other non-carbonaceous compounds present in the water sample. A condenser cools the gas stream to less than 50°C.

The CO_2 in the gas stream is measured by a non-dispersive infrared analyzer. The measured CO_2 is proportional to the Total Organic Carbon in the water sample. The price of this instrument is expected to be about $8,500.

Figure 15 is a flow diagram of the "Process Carbonaceous Analyzer" made by Beckman. This is another automated "on-stream" instrument that employs the combustion / infrared method of carbon analysis. However, it differs from the preceeding instrument in a significant way. The principle of operation is shown in this schematic drawing. Water from the sample source flows continuously through a 20 microliter hole in the slider insert of the valve. At intervals of from 2 to 5 minutes the slider operates to transfer this measured sample into an air carrier stream. The air carrier injects the sample into a catalytic combustion furnace maintained at 950°C. Here the water is vaporized and the carbonaceous material is oxidized to carbon dioxide. The air stream then carries the steam and CO_2 out of the furnace where the steam is condensed and removed. The CO_2 is measured in a non-dispersive infrared analyzer. The important difference between this and the preceding instrument is that this one also measures the inorganic carbon in the sample. Another difference is that with this instrument, it is recommended that the sample be filtered because the tolerance for solids has not yet been established. The price of this instrument is about $9,200.

Another similar instrument is Union Carbide Corporation's "Total Carbon Analyzer" shown schematically in Figure 16. It is quite similar to the preceding instrument, except that an inert gas, nitrogen, is used as the carrier gas. The sample can be injected manually or the instrument will work automatically for 24 hours.

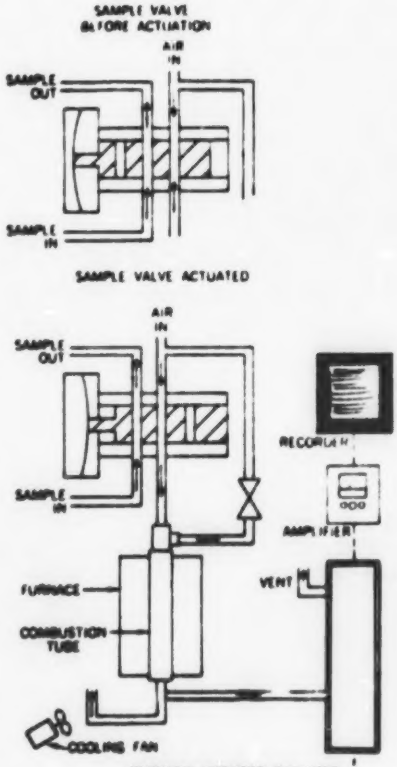


FIG.15

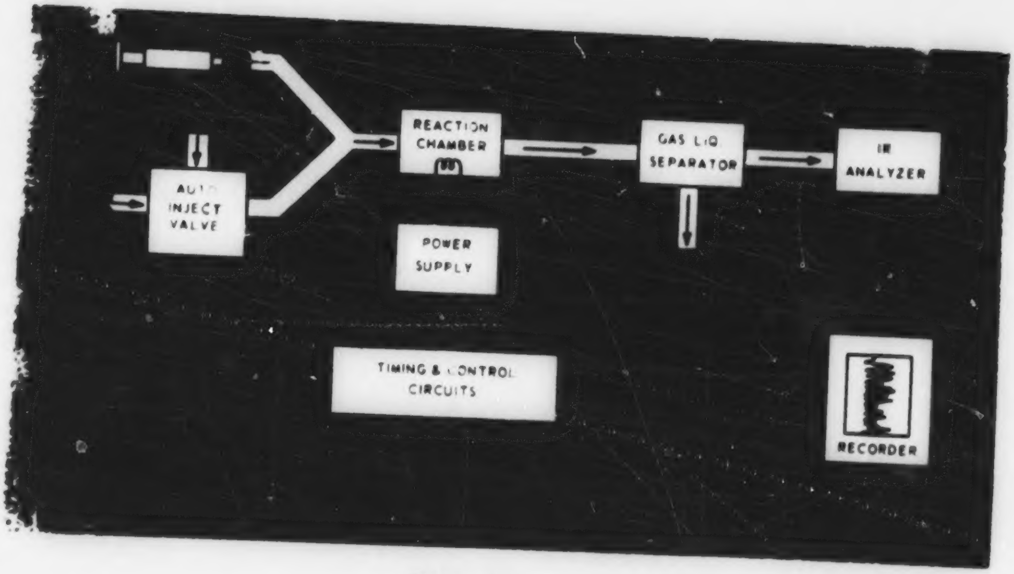

AUTO
INJECT
VALVE
REACTION
CHAMBER
GAS LIQ
SEPARATOR
IR
ANALYZER
POWER
SUPPLY
TIMING & CONTROL
CIRCUITS
RECORDER

FIG. 16

This outline was prepared by W. McMichael, Mechanical Engineer, Advanced Waste Treatment Laboratory, OWP, EPA, Cincinnati, OH 45226.

SECTION III IMPROVING CONVENTIONAL PLANT PERFORMANCE

	Outline Number
Case Histories: Effluent Excellence from Presently Available Secondary Treatment Processes	15
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Sludge Disposal by Landspreading	24

CASE HISTORIES: EFFLUENT EXCELLENCE FROM PRESENTLY AVAILABLE SECONDARY TREATMENT PROCESSES

I INTRODUCTION

A Pollution can be greatly reduced, almost overnight, by maximum use of existing secondary treatment processes.

B Activated sludge plants can produce final effluents containing considerably less than 10 mg/l suspended solids and 5-day BOD. Overall reductions of 95 to 99 percent are possible.

To obtain such results; the plants must be properly designed with adequate built in capacity and flexibility; plant characteristics must be appropriate to the incoming load; and the process must be skillfully controlled by conscientious qualified operators.

II SIOUX FALLS, SOUTH DAKOTA

A Plant Description

Treats 3.5 mgd strong meat-packing waste and 6.0 mgd domestic sewage. High rate trickling filters pre-treat packing plant wastes; activated sludge treats domestic sewage and polishes industrial wastes.

B Feature Story - 99% Reductions

Has demonstrated ability to provide 99% reductions during summertime (August) when plant was operating within design loading (Figures 1 and 2). BOD reduced from 900 to 9; TSS from 650 to 5 (Figures 3 and 4).

C Improved Operational Control

Wintertime pollutional load to river cut in half by improved operational control alone.

1 Control tests

Started using:

Settlometers for sludge density

Centrifuge for sludge density

Blanket finder for clarifier sludge blanket

Turbidimeter for current effluent quality

2 Interpretation

Sludge condition, system equilibrium, and process demands are the key items for system control.

3 Control adjustments

a Increased return sludge percentage from 30% to more than 100%.

b "Tight-rope" sludge wasting control to increase sludge concentration and activity without upsetting aerators and clarifiers.

4 Improved effluent quality

Former 30 mg/l BOD reduced to 20 mg/l

Former 35 mg/l TSS reduced to 13 mg/l

D Significant Loading Characteristics

See Table No. 1.

E Favorable Features at the Existing Plant:

1 Dedicated supervision and operation

2 Efficient aerators (Not spiral flow)

3 Effective suction type final clarifier sludge removal mechanism

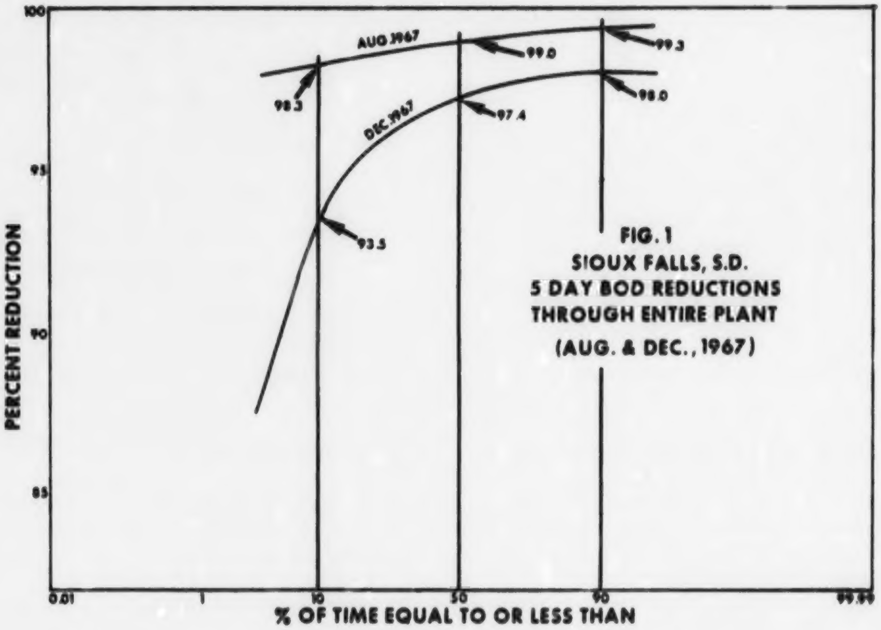

FIG. 1
SIOUX FALLS, S.D.
5 DAY BOD REDUCTIONS
THROUGH ENTIRE PLANT
(AUG. & DEC., 1967)

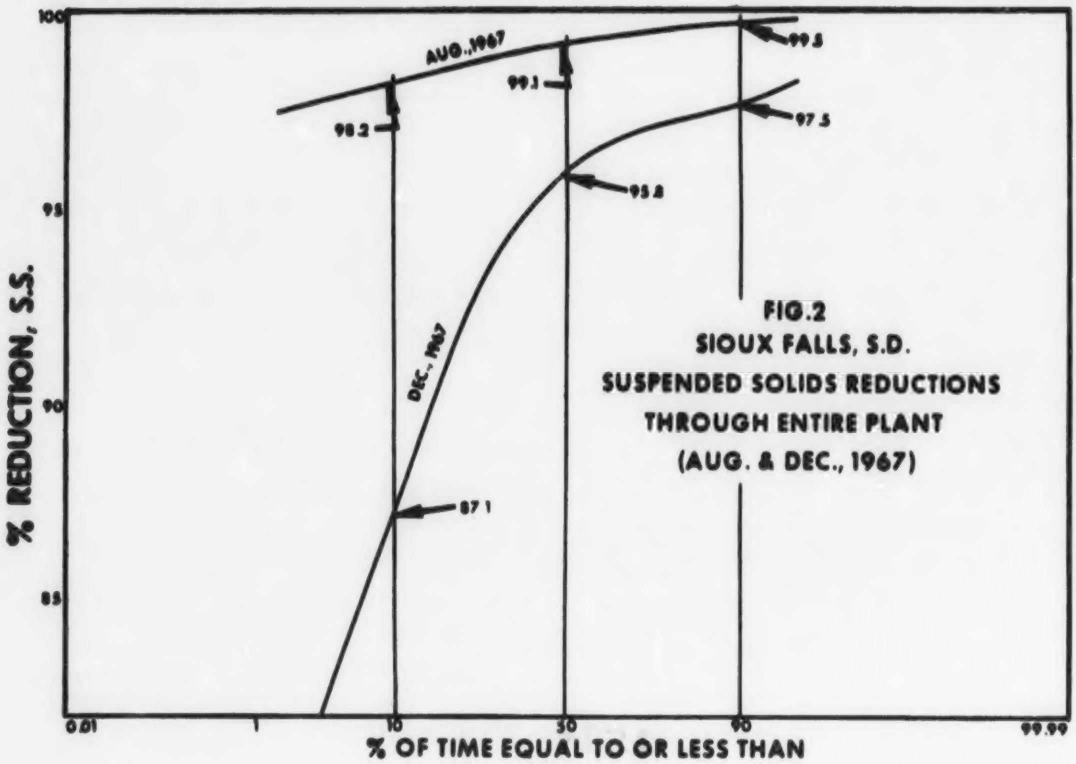

FIG.2
SIOUX FALLS, S.D.
SUSPENDED SOLIDS REDUCTIONS
THROUGH ENTIRE PLANT
(AUG. & DEC., 1967)

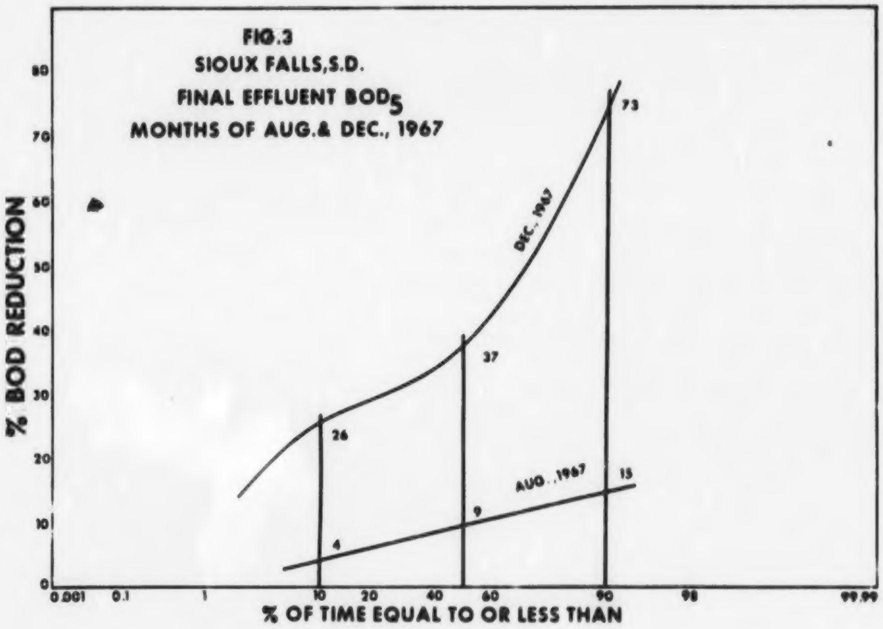

FIG.3
SIOUX FALLS,S.D.
FINAL EFFLUENT BOD_5
MONTHS OF AUG.& DEC., 1967

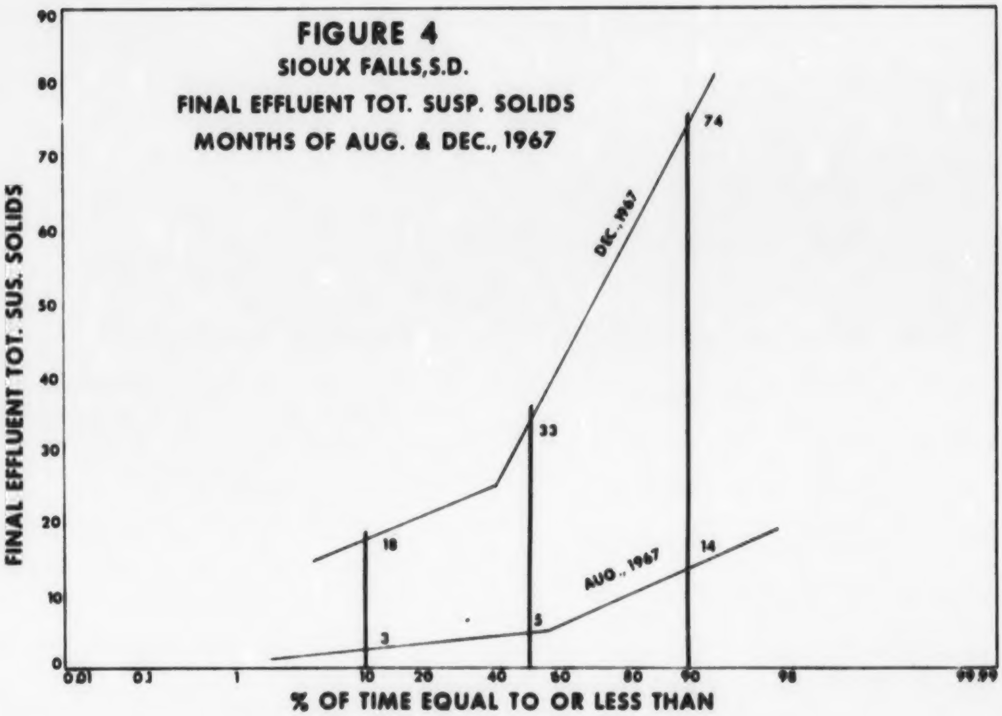

FIGURE 4
SIOUX FALLS,S.D.
FINAL EFFLUENT TOT. SUSP. SOLIDS
MONTHS OF AUG. & DEC., 1967

TABLE NO. 1

SIGNIFICANT LOADING CHARACTERISTICS
(Sioux Falls, South Dakota - Activated Sludge Plant)

	SUMMER	WINTER
BOD Load to Aerators		
Pounds per day	18,000	37,400
Pounds per 1,000 cu.ft. Aerator	103	160
Pounds per pound mixed liquor solids	1.3	1.2
Clarifier Surface Loading Rate		
Gals./sq.ft./day	720	640
BOD Reductions		
Total - Trickling filter & activated sludge	99	97
Activated sludge alone	95	90
TSS Reductions		
Total - Trickling filter & activated sludge	99	96
Activated sludge alone	96	83

Air (Approximate Range)

12 to 36 million cu.ft./day
0.5 to 3.0 cu.ft./gal.
300-1,000 cu.ft./lb. BOD

Return Sludge - 30% to 200%

Aeration Detention Time - 1.5 to 3.0 hr.

4 Complete mixed aerator flow pattern

5 Adequate return sludge pumping capacity

F Plant Deficiencies

1 Activated sludge plant overloaded

2 Trickling filters freeze up in winter

3 Disproportionately small aerator/clarifier volume

4 Not enough air

5 No scum removers on final clarifiers

6 Impossible to make precise control adjustments. (Not enough meters, no remotely controlled mechanical valve actuaters, and no automatic sensor-controllers.)

III METROPOLITAN ST. LOUIS SEWER DISTRICT - MSD

Coldwater Creek Wastewater Treatment Plant (Activated Sludge)

A Basically, this is a conventional 21 mgd standard rate activated sludge plant.

There are 6 aerators with spiral flow pattern diffusers and 4 final clarifiers with plow-type sludge scrapers.

B Feature Story - Pollution reduced to 1/4 its former strength by operation control alone.

1 Suspended Solids

		Before	After
Raw	(mg/l)	173	198
Primary Effluent	(mg/l)	155	142
Final Effluent	(mg/l)	92	16
Activated Sludge Reduction		40%	89%
Total Plant Reduction		46%	92%

2 5-Day BOD

		Before	After
Raw	(mg/l)	150	162
Primary Effluent	(mg/l)	152	130
Final Effluent	(mg/l)	40	9
Activated Sludge Reduction		74%	93%
Total Plant Reduction		73%	94%

D Improved Operational Control

1 Changed aerator/clarifier ratio: - by taking one of the 4 clarifiers out of service and placing an additional aerator in service. Theoretical full load characteristics as follows:

	Former	Changed To
Aerator/Clarifier Combination	3A/4C	4A/3C
Mixed Liquor (% by Centrifuge)	7.2	3.4
Return Sludge (% of Sewage Flow)	93	30
Flow Capacity (MGD)	13.2	14.5
Aerator Detention (HRS)	3.0	5.4
Clarifier Detention (HRS)	2.0	2.0

2 Control Tests & Interpretation

Introduced use of settlometers, centrifuge, blanket finder and turbidimeter to detect process balance and demands.

3 Control Adjustments

In this case: Reduced return sludge pumping. Increased air supply and excess sludge-wasting rates.

E Favorable Features at the Existing Plant

1 Dedicated supervision and operation

2 Adequate plant capacity for present dry weather flow

3 Multiple unit aerator and clarifier groups

F Plant Deficiencies

1 Plow-type sludge scrapers in final clarifiers

2 Spiral flow diffuser placement in aerators

3 Hydraulic short circuiting plus strong velocity currents in final clarifiers

4 No scum removal for final clarifiers

5 Inadequate return sludge capacity

6 Meter problems, and lack of remotely controlled mechanical valve actuators or automatic sensor-controllers limit process controlability.

G Present Status

1 Presently obtaining 90% reductions

2 Could cut pollutional load in half again if obvious plant deficiencies were corrected

IV CONCLUSIONS

A The activated sludge process can produce sparkling clear final effluents.

Plants can and should be properly designed (or modified) and operated to obtain their maximum inherent purification efficiencies needed to abate pollution of our national water resources. These include:

B Design

1 Provide adequate capacity for growth and equipment outages.

2 Flexibility - (Give the operators a chance.)

Provide ability to increase or decrease number of aerators or clarifiers, convert to a complete mixed system or to step aeration as required.

3 Avoid "spiral flow" aeration.

4 Use suction devices in final clarifiers for rapid removal of fresh sludge. Also provide surface scum removers.

5 Make air supply, sludge return and sludge wasting equipment truly variable and conveniently controllable.

6 Provide essential meters and sensors, remote valve actuators and automatic ratio controllers where required.

C Operation

1 Recruit and retain conscientious, intelligent, trained, and certified plant operators.

2 Provide practical "on-the-job" work experience type training.

3 Provide 24-hour around-the-clock operation. Test, evaluate and adjust process at least once per 8-hour shift.

4 Make the best use of existing facilities, guide counsel, and consulting engineers in designing needed improvements and additions.

This outline was prepared by A.W. West, Chief, Operation and Design, Division of Field Investigations, Cincinnati Center, EPA, Cincinnati, OH 45268.

CHAPTER 13

CASE HISTORIES OF TREATMENT PLANT UPGRADING

13.1 General

The capabilities and limitations of various unit processes have been discussed in preceding chapters. However, in any wastewater treatment plant, the operation of an individual process affects the operation of other processes in the treatment system. Therefore in upgrading situations, emphasis should be placed on the effects of upgrading on the treatment plant as a whole rather than on a particular unit operation.

Various case histories are discussed to illustrate procedures which have been used to upgrade wastewater treatment facilities. Cost estimates for these cases are furnished when such information is available.

13.2 Case History No. 1

Upgrading Using Chemical Addition to Primary Clarifiers and Conversion of Conventional Activated Sludge to Contact Stabilization (1)

Plant 1 before upgrading was a parallel activated sludge and trickling filter plant. The original nominal design capacity of the plant was 3.6 mgd, but the average influent flow had increased to 6.0 mgd. The plant was removing significantly less than 90 percent of the BOD during the summer and as little as 60 percent during the winter.

Upgrading of this plant was an interim measure, since the plant was scheduled to be replaced in 5 years with a regional system. Figure 13-1 is a flow diagram of the upgraded treatment system. Table 13-1 summarizes plant operating conditions prior to upgrading and plant design conditions after upgrading. The conventional activated sludge system originally designed to treat a flow of 1.2 mgd, was upgraded to a contact stabilization system. This was accomplished by increasing the capacity of the mechanical aerators from a total of 40 hp up to 110 hp (60 hp in the contact basin and 50 hp in the stabilization basin) and by modifying the basin's piping. Since experience has shown that contact stabilization works well without primary clarification, the existing primary clarifier was converted to a secondary clarifier to decrease the activated sludge secondary clarifier overflow rate. The contact stabilization process was designed to handle a flow of 3.0 mgd.

The trickling filter portion of the plant was originally designed to operate at 2.4 mgd as a two-stage system. However, during periods of hydraulic overloading, the trickling filters were operated in parallel to increase the hydraulic capacity.

Plant operating data indicated that the secondary portion of the parallel-operated trickling filter system was removing 70-80 percent of the primary effluent BOD. Therefore, it was

This copy taken from the "Process Design Manual for Upgrading Existing Wastewater Treatment Plants" prepared by Roy F. Weston, Inc. under Contract 14-12-933 with the EPA Technology Transfer Program.

FIGURE 13-1
CASE HISTORY NO. 1
FLOW DIAGRAM OF PLANT INCLUDING UPGRADED UNIT PROCESSES

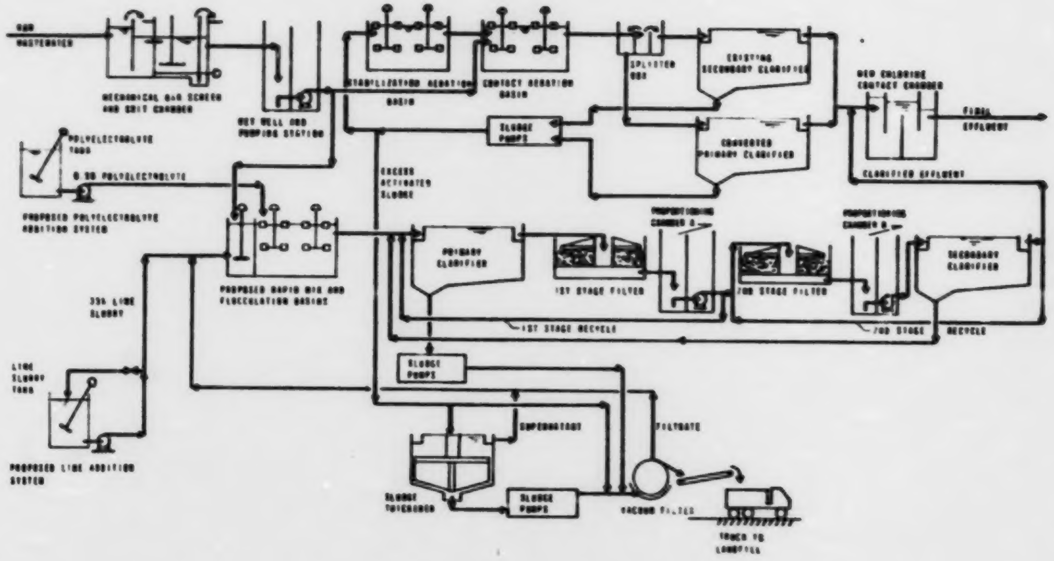
RAW WASTEWATER
MECHANICAL BAR SCREEN AND GRIT CHAMBER
WET WELL AND PUMPING STATION
STABILIZATION AERATION BASIN
CONTACT AERATION BASIN
SPLITTER BOX
EXISTING SECONDARY CLARIFIER
CONVERTED PRIMARY CLARIFIER
SLUDGE PUMPS
NEW CHLORINE CONTACT CHAMBER
FINAL EFFLUENT
CLARIFIED EFFLUENT
POLYELECTROLYTE TANK
0.5% POLYELECTROLYTE
PROPOSED POLYELECTROLYTE ADDITION SYSTEM
EXCESS ACTIVATED SLUDGE
PROPOSED RAPID MIX AND FLOCCULATION BASINS
PRIMARY CLARIFIER
1ST STAGE FILTER
PROPORTIONING CHAMBER 1
2ND STAGE FILTER
PROPORTIONING CHAMBER 2
SECONDARY CLARIFIER
1ST STAGE RECYCLE
2ND STAGE RECYCLE
33% LIME SLURRY
LIME SLURRY TANK
PROPOSED LIME ADDITION SYSTEM
SLUDGE PUMPS
SUPERNATANT
FILTRATE
SLUDGE THICKENER
SLUDGE PUMPS
VACUUM FILTER
TRUCK TO LANDFILL

Table 13-1

Case History No. 1 - Plant Operating and Design Conditions

Operation	Operating Conditions Before Upgrading	Design Conditions After Upgrading	Remarks
Pre-treatment Facilities (Bar Screen, Aerated Grit Chamber)			
Average Flow, mgd	6.0	6.0	
Activated Sludge Plant			
Primary Clarifier Overflow Rate, gpd/sq.ft.[1]	1,200	-	Primary clarifier converted to secondary clarifier
Flow to Aeration Basin, mgd	3.0	3.0	
Volumetric Loading in Aeration Tank, lbs. BOD/day/1,000 cu.ft	-	60[2]	O$_2$ supply, not organic loading, was limiting prior to upgrading.
Average Sludge Recycle, percent	30	100	
Secondary Clarifier Overflow Rate, gpd/sq.ft.[1]	960	740	Reduction in overflow rate is achieved by converting the primary clarifier to a secondary unit.
BOD Removal in Secondary Treatment, percent	75	90	
Suspended Solids Removal in Secondary Treatment, percent	75	90	
Trickling Filter Plant			
Primary Clarifier Overflow Rate, gpd/sq.ft.[1]	820	820	
BOD Removal in Primary Treatment, percent	20	50 to 60	Chemical coagulation
Flow to Trickling Filter, mgd	3.0	3.0	
Hydraulic Loading, mgd/acre-ft.[3]	-	6.25	For each stage
Organic Loading, lbs. BOD/day/1,000 cu.ft.[3]	-	72	First Stage
		19	Second Stage
Recirculation Ratio	1.0	1.0	For each stage
Secondary Clarifier Overflow Rate, gpd/sq.ft.[4]	1,000	1,000	
BOD Removal in Secondary Treatment, percent	70 to 80	70 to 80	
Sludge Handling Facilities			
Vacuum Filter Operation	16 hrs./day 5 days/week	20 hrs./day 6 days/week	Increased operation to accommodate increased sludge quantities

[1] Clarifier overflow rates based on average flow.

[2] Organic loading increases due to increase in flow and elimination of primary treatment.

[3] Does not include recirculation.

[4] Based on average flow plus recycle.

decided to confine the upgrading to improvement of the removal efficiency of the primary clarifier. The selected approach was chemical addition, namely lime and anionic polyelectrolyte. Laboratory testing indicated that as much as 60 percent BOD removal and 75 percent suspended solids removal could be attained in the primary clarifier with the addition of 1.0 mg/l polyelectrolyte and 200 mg/l of lime. With this modification, the trickling filter system could again be operated as a 2-stage system at a flow of 3.0 mgd.

A summary of the measured performance of the existing system and the anticipated performance after upgrading is given below in Table 13-2.

Table 13-2

Summary of Treatment Performance
for Case History No. 1

	Measured Performance Before Upgrading	Anticipated Performance After Upgrading
Design Flow, mgd	3.6	6.0
Average Flow, mgd	6.0	6.0
BOD Removal, percent	80	90
Effluent BOD, mg/l	40	20
SS Removal, percent	85	90
Effluent SS, mg/l	30	20

The cost of upgrading the plant was estimated at $510,000, and the increased annual operating costs were estimated at $74,800. These costs are broken down in Table 13-3.

Table 13-3

Summary of Upgrading Costs
for Case History No. 1
(ENR Index 1500)

ESTIMATED CONSTRUCTION COST

Chemical Addition/Flocculation	$ 45,000
Conversion to Contact Stabilization	340,000
Mechanical Refurbishing	25,000
Electrical Refurbishing	50,000
Construction Contingency	50,000
Total Estimated Construction Cost[1]	$510,000
Unit Capital Cost of Upgrading	$212/1,000 gpd of upgraded flow

ESTIMATED ANNUAL OPERATING COST INCREASE

Operation and Maintenance	$ 5,700
Electrical	9,100
Chemical	52,000
Sludge Disposal	8,000
Total Annual Operating Cost Increase	$ 74,800
Unit Annual Operating Cost Increase	3.4 cents/1,000 gallons of treated flow

[1] Costs do not include engineering, legal, and administrative fees.

13.3 Case History No. 2

Upgrading an Existing High-Rate Trickling Filter Plant Using a Series Activated Sludge Process (2)

Plant 2 before upgrading was a high-rate trickling filter plant whose original design capacity of 12.5 mgd was being approached. The initial treatment requirements were 85 percent BOD and SS removals, and removals of approximately 80 percent were actually being obtained. The treatment requirements were then changed to include 90-95 percent removal of the total oxygen demand (the sum of the ultimate carbonaceous oxygen demand and the nitrogenous oxygen demand). This additional requirement indicated that most of the ammonia nitrogen had to be removed or converted to the nitrate form.

Figure 13-2 is a flow diagram of the upgraded treatment system. Table 13-4 summarizes plant operating conditions prior to upgrading and plant design conditions after upgrading. An in-plant survey indicated that the existing equipment was in good condition and could be incorporated into the upgrading scheme. After an evaluation of alternative upgrading possibilities, the existing plant was upgraded from 12.5 mgd to 22.5 mgd using a series treatment of activated sludge and trickling filtration. The activated sludge system was designed with a detention time of 3 hours, with the activated sludge effluent being subsequently treated in the trickling filter to take advantage of the incipient nitrifying ability of the trickling filter.

Another major consideration in the upgrading was the disposal of waste activated sludge. To reduce the volume of primary and waste activated sludge, a gravity thickener was incorporated into the design. Although the existing two-stage anaerobic digesters had sufficient volume to handle the thickened sludge from the upgraded plant, operational problems were being experienced in the existing plant due to incomplete mixing and non-uniform temperature control in the primary digesters. The original design included mechanical mixing using draft tubes and internal heating coils. To improve digester performance, gas recirculation was used for better mixing, and an external heat exchanger was incorporated to facilitate maintenance of a uniform temperature throughout the primary digesters. When the upgraded plant comes on line, it is expected to perform as indicated in Table 13-5.

FIGURE 13-2
CASE HISTORY NO. 2
FLOW DIAGRAM OF PLANT INCLUDING UPGRADED UNIT PROCESSES

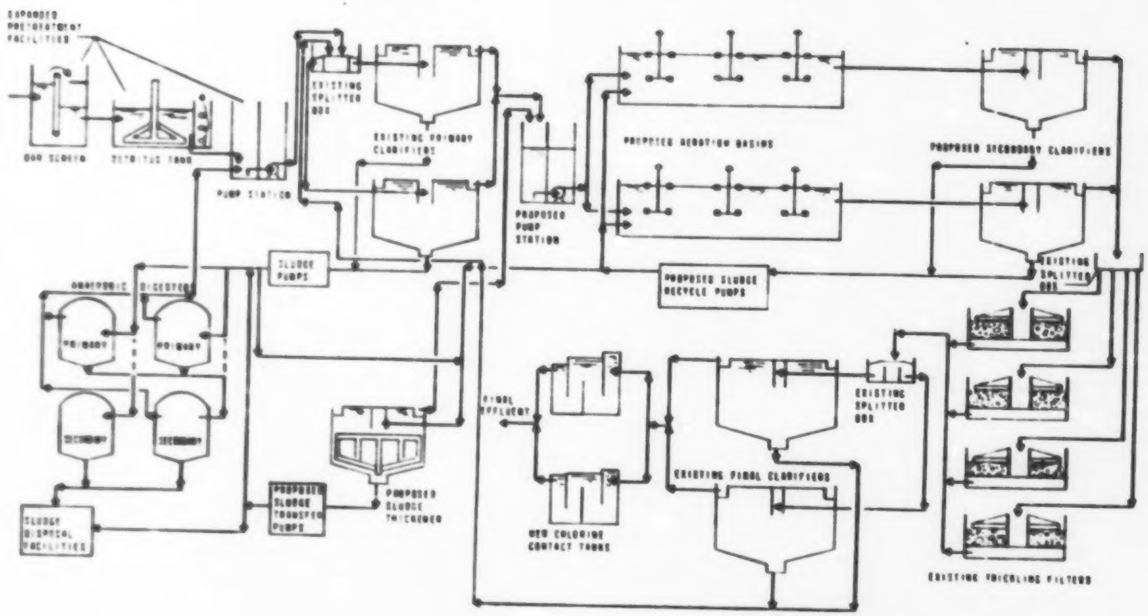
EXPANDED PRETREATMENT FACILITIES
BAR SCREEN
DETRITUS TANK
PUMP STATION
EXISTING SPLITTER BOX
EXISTING PRIMARY CLARIFIERS
PROPOSED PUMP STATION
PROPOSED AERATION BASINS
PROPOSED SECONDARY CLARIFIERS
EXISTING SPLITTER BOX
PROPOSED SLUDGE RECYCLE PUMPS
SLUDGE PUMPS
ANAEROBIC DIGESTERS
PRIMARY
PRIMARY
SECONDARY
SECONDARY
SLUDGE DISPOSAL FACILITIES
PROPOSED SLUDGE TRANSFER PUMPS
PROPOSED SLUDGE THICKENER
FINAL EFFLUENT
NEW CHLORINE CONTACT TANKS
EXISTING FINAL CLARIFIERS
EXISTING SPLITTER BOX
EXISTING TRICKLING FILTERS

Table 13-4

Case History No. 2 - Plant Operating and Design Conditions

Operation	Operating Conditions Before Upgrading	Design Conditions After Upgrading
Pre-treatment Facilities (Bar Screen, Grit Chamber, Raw-Sewage Pump Station) Avg. Capacity, mgd	12.5	22.5
Primary Clarifiers		
Overflow Rate, gpd/sq.ft.	766[1]	785[2]
BOD Removal, percent	56	30
SS Removal, percent	74	60
Aeration Basins		
Detention Time, hours	–	3.0
Volumetric Loading, lbs. BOD/day/ 1,000 cu.ft.	–	30
Organic Loading, lbs. BOD/day/lb. MLVSS	–	0.25
Sludge Recycle, percent of average flow	–	60
Mechanical Aerator Capacity, hp	–	750
Intermediate Clarifiers		
Overflow Rate, gpd/sq.ft.		785
Trickling Filter		
Hydraulic Loading, mgad	30	21.6
Organic Loading, lbs. BOD/day/ 1,000 cu.ft.	66	21
Recirculation Ratio	1.5[3]	0
Final Clarifiers		
Overflow Rate, gpd/sq.ft.	766[4]	785
Sludge Thickener		
Design Underflow Concentration	–	4.0
Design Solids Loading, lbs. SS/day/sq.ft.	–	10.0
Detention Time, hours	–	10.0
Anaerobic Digesters		
Loading, lbs. VSS/day/cu.ft.	0.05	0.12
Detention Time, days	34	14

[1] Based on average design flow plus 75% secondary sludge recirculation.

[2] Excluding sludge recycle.

[3] The recirculation ratio includes 75% secondary sludge recirculation to the primary clarifier influent and 75% secondary clarifier effluent recirculation to the trickling filter influent.

[4] Based on average design flow plus 75% secondary clarifier effluent recirculation.

Table 13-5

Summary of Treatment Performance for Case History No. 2

	Before Upgrading	After Upgrading
Average Design Flow, mgd	12.5	22.5
BOD Removal, percent	81	>90
Effluent BOD, mg/l	41	15
SS Removal, percent	81	>90
Effluent SS, mg/l	40	<10
Effluent NH_3-N, mg/l	27	2
Effluent NO_3-N[1], mg/l	<1	26

[1] Approximate

The construction costs for treatment facilities were estimated at $4,130,000, and the increased annual operating costs due to upgrading were estimated at $224,000. These costs are broken down in Table 13-6.

Table 13-6

Summary of Upgrading Costs for Case History No. 2 (ENR Index 1500)

ESTIMATED CONSTRUCTION COST

Expansion of Pre-treatment Facilities	$ 165,000
Expansion of Primary Treatment	190,000
Addition of Activated Sludge Treatment and Intermediate Clarification	2,030,000
Expansion of Sludge Handling	54,000
Expansion of Control Building, Chlorination Facilities, etc.	260,000
Piping, Electrical, Instrumentation, etc.	890,000
Construction Contingency	541,000
Total Estimated Construction Cost[1]	$4,130,000
Unit Capital Cost of Upgrading	$413/1,000 gpd of upgraded flow

ESTIMATED ANNUAL OPERATING COST INCREASE

Labor	$ 30,000
Power	86,000
Maintenance and Supplies	108,000
Total Annual Operating Cost Increase	$224,000
Unit Annual Operating Cost Increase	2.7 cents/1,000 gallons of treated flow

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[1] Costs do not include engineering, legal, and administrative fees.

13.4 Case History No. 3

Use of Roughing Filter to Upgrade an Existing Low-Rate Trickling Filter Plant (3) (4 (5)

Case History No. 3 involves the upgrading of an existing low rate trickling filter plant in Huber Heights, Ohio. The original plant was designed in August, 1956 for a flow of 0.7 mgd, with 85 percent BOD and suspended solids removals. The community developed so rapidly that by 1970, the average flow had increased to 2.3 mgd. The flow diagrams for the original and upgraded plant are shown in Figure 13-3.

Operational and performance data for the overloaded plant for 1962, when the plant was receiving 1.15 mgd, are compared with corresponding data after the plant was upgraded to 2.3 mgd in Table 13-7.

The comminuter and primary clarifiers were replaced with three Hydrasieve units of 1-mgd capacity each. These units are stationary screens capable of removing 20 to 35 percent of the BOD and suspended solids. A Hydrasieve unit is illustrated in Figure 13-4, along with a schematic flow diagram through the unit. These screens generally require no power and little maintenance.

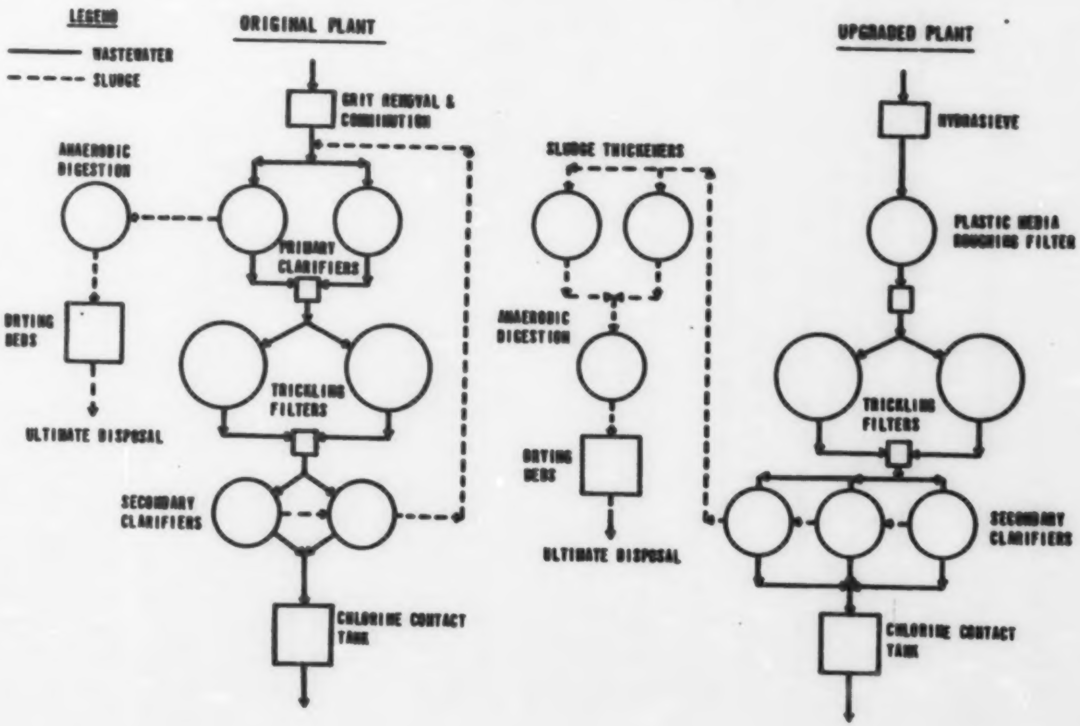
The plastic media roughing filter used in the upgrading has an application rate of approximately 2.5 gpm/sq.ft. Present BOD removal is about 25 to 35 percent through the roughing unit. Because of the increased hydraulic loading, it was necessary to expand the secondary clarification and chlorine contact tank capacities. The abandoned primary clarifiers were converted into sludge thickeners. This step, in addition to conversion of the anaerobic digester to a high-rate unit using gas recirculation for mixing, enabled the sludge handling system to process the increased quantity of sludge produced.

This case history points out the fact that an existing plant may be gradually upgraded to handle a three-fold increase in flow with the use of innovative techniques and newly applied process equipment. The capital costs for upgrading the capacity of the plant were estimated at approximately $300,000 (ENR 1500), or $187 per 1,000 gallons/day of upgraded flow capacity.

FIGURE 13-3

CASE HISTORY NO.3

COMPARISON OF ORIGINAL AND UPGRADED FLOW DIAGRAMS


LEGEND
WASTEWATER
SLUDGE
ORIGINAL PLANT
GRIT REMOVAL & COMMINUTION
ANAEROBIC DIGESTION
PRIMARY CLARIFIERS
DRYING BEDS
TRICKLING FILTERS
ULTIMATE DISPOSAL
SECONDARY CLARIFIERS
CHLORINE CONTACT TANK
SLUDGE THICKENERS
ANAEROBIC DIGESTION
DRYING BEDS
ULTIMATE DISPOSAL
UPGRADED PLANT
HYDRASIEVE
PLASTIC MEDIA ROUGHING FILTER
TRICKLING FILTERS
SECONDARY CLARIFIERS
CHLORINE CONTACT TANK

171<

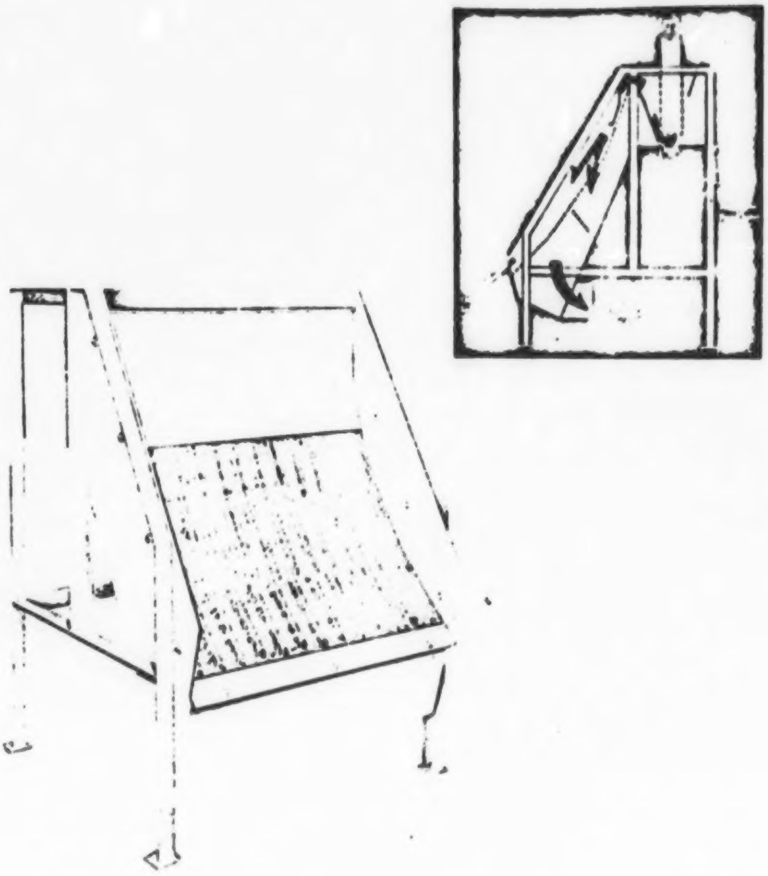
Table 13-7

Case History No. 3 - Plant Operating and Performance Data

Operation	Before Upgrading (1962)	After Upgrading
Average Flow Rate, mgd	1.15	2.3
Primary Clarifier		
Overflow Rate[1], gpd/sq.ft.	1,170	–[5]
Hydrasieve Slot Size, inches	–	0.06
BOD Removal, percent	35[2]	25[3]
SS Removal, percent	–[6]	22[3]
Plastic Media Roughing Filter		
Hydraulic Loading[4], gpm/sq.ft.	–	2.5
Organic Loading[4], lbs. BOD/day/1,000 cu.ft.	–	520
Recirculation Ratio	–	≈2.0
BOD Removal, percent	–	30
Trickling Filter (Stone Media)		
Hydraulic Loading, mgad	6.0	12.0
Organic Loading, lbs. BOD/day/1,000 cu.ft.	56.2	87.0
Secondary Clarifiers		
Overflow Rate, gpd/sq.ft.	1,170	750
Overall Plant Performance		
BOD Removal, percent	83	85
Effluent BOD, mg/l	41	37
SS Removal, percent	–[6]	84
Effluent SS, mg/l	–[6]	40

[1] Based on average flow rate.
[2] Based on primary clarifier performance.
[3] Based on hydrasieve performance only.
[4] Including recirculation.
[5] Primary clarifiers converted to gravity thickeners.
[6] Operating data not available.

FIGURE 13.4
HYDRASIEVE SCREENING UNIT*



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*COURTESY OF THE BAUER BROS. CO. - SPRINGFIELD, OHIO

13.5 Case History No. 4

Upgrading an Existing High-Rate Trickling Filter by Conversion to a Super-Rate Filter System (6)

The North Treatment Plant at Sedalia, Missouri, is a high-rate trickling filter plant, designed for 1.25 mgd, and was removing 85 percent of the BOD in 1963. However, the Water Pollution Board set a final effluent BOD of 20 mg/l, which Sedalia could not meet with the existing facilities. The 1963 plant flow diagram is illustrated in Figure 13-5.

The plant was upgraded to treat an average design flow of 2.5 mgd. The existing stone-media filter was renovated to operate with plastic media. In addition, a second plastic-media filter was constructed. The two plastic-media filters are operated in parallel with a total recirculation ratio of 1.55. One additional primary clarifier and one additional secondary clarifier were installed.

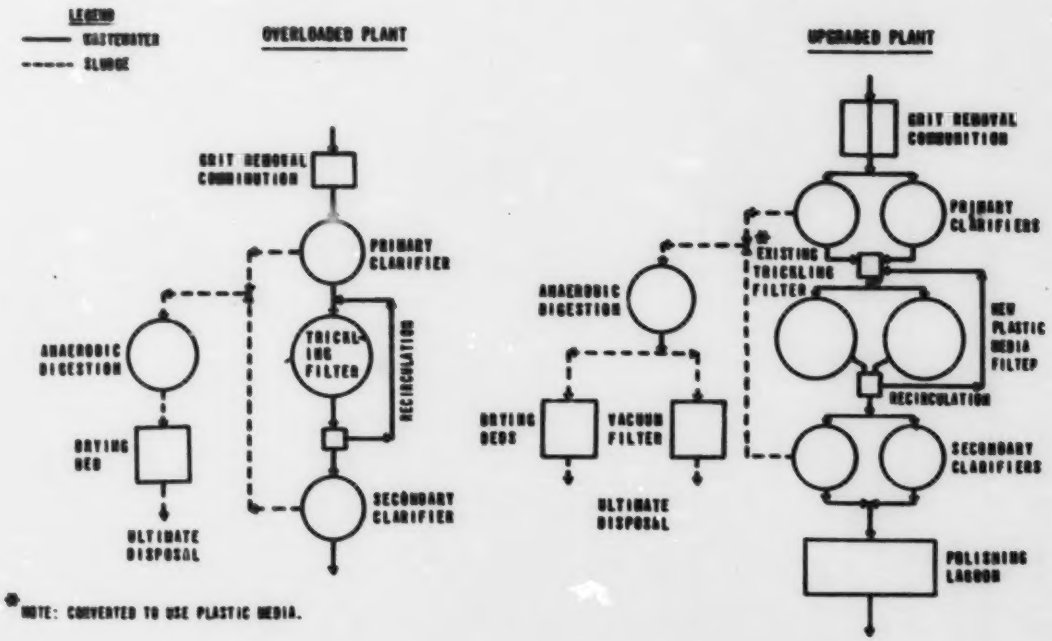
To remove additional BOD and suspended solids, a shallow aerobic polishing lagoon was constructed after the secondary clarifiers, and a vacuum filter was added to reduce the volume of digested sludge. A flow diagram of the upgraded plant is also shown in Figure 13-5. Table 13-8 contains a summary of operating data for the 1963 overloaded period; in addition, the upgraded design criteria are listed, along with actual operational data for the post-upgrading period. It should be noted that the effluent BOD was improved from 115 mg/l to 11 mg/l after upgrading, which is below the 20 mg/l requirement. It should also be pointed out that Missouri has no suspended solids removal requirements for plants with a flow of less than 10 mgd.

The capital costs of upgrading the plant were estimated at $2,600,000 (ENR Index 1,500), or $2,080 per 1,000 gpd of incremental upgraded capacity.

FIGURE 13-5

CASE HISTORY NO. 4

COMPARISON OF ORIGINAL AND UPGRADED FLOW DIAGRAMS


LEGEND
WASTEWATER
SLUDGE
OVERLOADED PLANT
GRIT REMOVAL COMMINUTION
PRIMARY CLARIFIER
TRICKLING FILTER
RECIRCULATION
SECONDARY CLARIFIER
ANAEROBIC DIGESTION
DRYING BED
ULTIMATE DISPOSAL
UPGRADED PLANT
GRIT REMOVAL COMMINUTION
PRIMARY CLARIFIERS
EXISTING TRICKLING FILTER
NEW PLASTIC MEDIA FILTER
RECIRCULATION
SECONDARY CLARIFIERS
POLISHING LAGOON
ANAEROBIC DIGESTION
DRYING BEDS
VACUUM FILTER
ULTIMATE DISPOSAL

*NOTE: CONVERTED TO USE PLASTIC MEDIA.

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Table 13-8

Case History No. 4 - Plant Operating and Design Data

Operation	1963 Operating Data	Upgraded Design	1969 Operating Data
Average Daily Flow, gpm	1.25	2.5	1.80
Raw Wastewater BOD, mg/l	768	576	450
Primary Clarifiers			
Overflow Rate[1], gpd/sq.ft.	1,000	1,000	720
BOD Removal, percent	40	40	60
Trickling Filters			
Hydraulic Loading[2], mgad	24.5	31.8	25
Organic Loading[2], lbs. BOD/day/ 1,000 cu.ft.	226	72.6	20
Recirculation Ratio	1.0	1.55	1.55
Final Clarifiers			
Overflow Rate[1], gpd/sq.ft.	755	755	545
Secondary BOD Removal, percent	75	93.2	96.8
Polishing Lagoon (Shallow Aerobic)			
Maximum BOD Loading, lbs. BOD/ acre/day	–	68	30
BOD Removal, percent	–	12	54
Vacuum Filtration Rate, lbs./sq.ft./hour	–	5.0	–[3]
Overall Plant Performance			
BOD Removal, percent	85	96.5	97.7
Effluent BOD, mg/l	115	20	11

[1] Based on average daily flow.

[2] Including Recirculation

[3] Lack of sufficient operating data.

13.6 Case History No. 5

Upgrading a Contact Stabilization Package Plant to a "Modified" Completely-Mixed Flow Pattern (7)

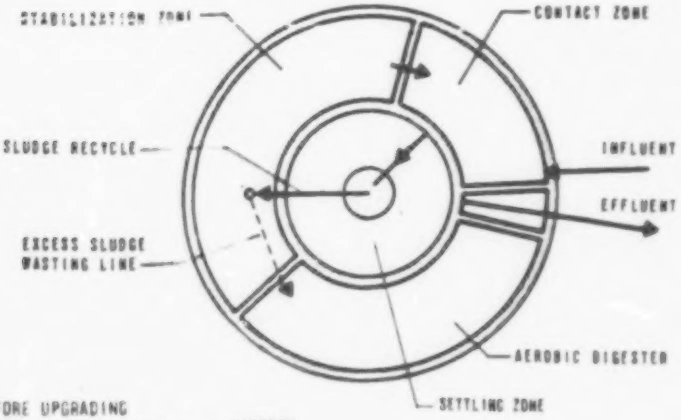
Coralville, Iowa, had a contact stabilization package plant which was providing detention times of 2.6 and 6.5 hours, respectively, in the contact and stabilization zones, based on a forward flow of 867,000 gallons per day, not including sludge recycle. A typical plan view of the contact stabilization package plant is shown in Figure 13-6.

As previously discussed in Chapter 5, a contact-zone detention time of this magnitude may result in improper plant operation because the sludge becomes partially stabilized in this zone and acquires poorer settling characteristics. After investigation, this was found to be the case in Coralville. Operational data from the plant before upgrading are summarized in Table 13-9. The effluent BOD and suspended solids were 26 and 24 mg/l, respectively.

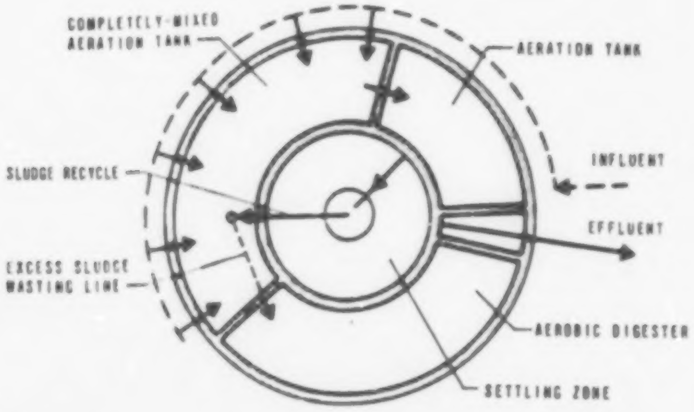
To improve the plant's performance, it was decided to modify the flow pattern as indicated in Figure 13-6. The influent piping was modified so that the raw wastewater was evenly distributed into what originally was the stabilization zone. No wastewater was introduced into the former contact zone. Mixed liquor in the upgraded system proceeded from the former stabilization zone through the former contact zone to the secondary clarifier. The return sludge was introduced into the former stabilization zone at one point only. Therefore, the upgrading resulted in a "modified" completely-mixed flow pattern, with an overall detention time of 9.1 hours for an average flow of 867,000 gpd.

Performance data for the upgraded plant are included in Table 13-9. The effluent BOD and suspended solids concentrations were lowered to 13 and 6 mg/l, respectively, by the upgrading procedure. The costs associated with this modification are primarily due to piping changes. No cost breakdown was available for the modification.

FIGURE 13-6
CASE HISTORY NO. 5
UPGRADING A CONTACT STABILIZATION PACKAGE PLANT
TO A COMPLETELY-MIXED FLOW PATTERN (5)


STABILIZATION ZONE
CONTACT ZONE
SLUDGE RECYCLE
INFLUENT
EFFLUENT
EXCESS SLUDGE WASTING LINE
AEROBIC DIGESTER
SETTLING ZONE

BEFORE UPGRADING
CONTACT STABILIZATION FLOW PATTERN


COMPLETELY-MIXED AERATION TANK
AERATION TANK
INFLUENT
SLUDGE RECYCLE
EFFLUENT
EXCESS SLUDGE WASTING LINE
AEROBIC DIGESTER
SETTLING ZONE

AFTER UPGRADING
MODIFIED COMPLETELY-MIXED FLOW PATTERN

Table 13-9

Case History No. 5 - Plant Operating Data

Operation	Contact Stabilization[1]	Modified Completely-Mixed[2]
Average Flow, mgd	0.867	0.867
Influent BOD, mg/l	135	135
Influent SS, mg/l	150	150
Aeration Tank		
Sludge Recycle, percent	60	60
Contact Unit Volumetric Load, lbs. BOD/day/1,000 cu.ft.	78	22[4]
Contact Unit Organic Load, lbs. BOD/day/lb. MLVSS	0.4	–
Contact Unit MLSS, mg/l	3,500	–
Secondary Clarifier		
Overflow Rate[3], gpd/sq.ft.	750	750
Overall Plant Performance		
BOD Removal, percent	81	90
Effluent BOD, mg/l	26	13
SS Removal, percent	84	96
Effluent SS, mg/l	24	6

[1] Before Upgrading
[2] After Upgrading
[3] Based on average flow
[4] Based on total aeration volume.

13.7 Case History No. 6

Upgrading Using Polyelectrolyte Addition Before the Primary Clarifier (8)

The Easterly Wastewater Treatment Works in Cleveland, Ohio, is a conventional activated sludge plant whose dry-weather design flow is 123 mgd. In 1968, the plant was hydraulically overloaded. Operational data from the period are presented in Table 13-10, identified as the 1968 Control Period. It is interesting to note that the overflow rate of the primary clarifiers was 2,030 gpd/sq.ft., which is substantially above the accepted values. As a result of the hydraulic overloading, the overall BOD and suspended solids removals were only 79 and 85 percent, respectively.

To improve the overall plant performance, a polyelectrolyte addition program was initiated. An anionic polyelectrolyte, Purifloc A-23, was added at an average dosage of 0.21 mg/l. Since proper polyelectrolyte dispersal and uniform mixing into the entire waste flow constitute an extremely important aspect of the flocculation process, it was decided to add the polymer at the plant's two Venturi meters. These meters are located immediately after the grit chamber and in front of the pre-aeration basin. Dye studies indicated that there was a 7.5-minute travel time between the Venturi meters and the primary clarifiers. Six of the 7.5 minutes would be spent in the pre-aeration basin. The gentle agitation in the pre-aeration basin provided adequate flocculation of wastewater solids.

A summary of the effectiveness of the polymer addition on plant performance as compared to performance during the previous control period is also presented in Table 13-10. The improvement in primary suspended solids removal increased the volume of primary sludge from 5.0 to 6.8 million gallons per month. Overall plant performance was improved, as noted by the reduced effluent BOD and suspended solids concentrations. In addition to the increased treatment efficiency, the polyelectrolyte addition was responsible for the following benefits to the subsequent activated sludge process:

1. A 20 percent volume decrease in the amount of waste activated sludge produced.
2. A 22 percent reduction in air supply requirements, resulting in a power cost savings of over $3,300 per month.
3. An increase in aeration tank D.O. concentration from an average of 3.2 mg/l to 3.8 mg/l.

An economic comparison was made between polyelectrolyte addition and providing additional tankage to equal the performance of the flocculation system. The amortized cost for the additional tankage was about $314,000 per year, while the chemical cost was $150,000 per year, thereby indicating a considerable yearly savings in favor of the polyelectrolyte alternative.

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Table 13-10

Case History No. 6 - Plant Operating Data

Operation	1968 Control Period	Polymer Demonstration Period
Influent BOD, mg/l	104	67
Influent SS, mg/l	169	157
Primary Clarifier		
Overflow Rate, gpd/sq.ft.	2,030	2,170
BOD Removal, percent	31	46
SS Removal, percent	31	51
Sludge Solids Concentration, percent	4.1	4.3
Sludge Volume Pumped, million gallons/month	5.0	6.8
Aeration Tank		
MLSS, mg/l	1,670	1,602
Organic Loading, lbs. BOD/day/lb. MLSS	0.48	0.29
Dissolved Oxygen, mg/l	3.2	3.8
Waste Activated Sludge Concentration, percent	2.4	2.0
Waste Activated Sludge Pumped, million gallons per month	12.3	9.8
Overall Plant Performance		
BOD Removal, percent	79.1	83.4
Effluent BOD, mg/l	21.8	11.1
SS Removal, percent	85.3	89.2
Effluent SS, mg/l	24.8	17.0

13.8 References

1. Roy F. Weston, Inc., Confidential Concept Engineering Report, April 6, 1970.

2. Roy F. Weston, Inc., Confidential Concept Engineering Report, April 2, 1970

3. Wittenmyer, J.D., and Sak, J.G., *Plastic Media Roughing Filter Provides Most Economical Plant Expansion*. Presented at the Ohio Water Pollution Control Association, June 15, 1967.

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5. Private communication with J.D. Wittenmyer, Vice-President, Ohio Suburban Water Company, Dayton, Ohio, January 22, 1971.

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7. Dague, R.R., et al, *Contact Stabilization Theory, Practice, Operational Problems and Plant Modifications*. Presented at the 43rd Annual Conference - WPCF, Boston, Mass. (October 4-9, 1970).

8. Wirts, J.J., *The Use of Organic Polyelectrolyte for Operational Improvement of Waste Treatment Processes*. Federal Water Pollution Control Administration, Grant No. WPRD 102-01-68, May, 1969.

REMOVAL OF PHOSPHORUS AND COLLOIDAL SOLIDS BY COAGULATION IN CONVENTIONAL TREATMENT

I INTRODUCTION

A Why are we here today?

1 Obviously any operation can be improved with better attention, understanding and input effort. This is particularly applicable in many wastewater treatment plants.

2 These plants are faced with effluent upgrading requirements, tired equipment, increased and more difficult loading. This calls for more effective use of what we have on a day-by-day basis.

3 Recent reports(1, 2, 3, 4, 5) show that progress has occurred in applying additional known techniques in existing facilities to improve cost-benefit ratios.

4 Chemical coagulation has almost universal application in water supply. Why not in wastewater treatment?

5 Coagulation can provide more days of better effluents to gain time and needed facilities as a trade off for chemical and control cost.

6 Dawn comes up slowly at the treatment works. This is not only because the plant usually is located at the lowest elevation. Mental fog sometimes is more difficult to dispel. An aggressive approach with positive pragmatism is required.

7 Designers must consider modification of existing facilities and provision for coagulation in new facilities. Designers should be part of start-up and initial operations until the personnel and facility function smoothly. The designer and operational control personnel must be a working team for mutual benefit.

B Should you use these processes? When and Where?

1 This is not intended as part of the great phosphorus debate. Look at your local Standards. BOD and solids are inseparable. Phosphorus cannot be removed without efficient solids removal.

2 Adapting existing plants for coagulation usually is simple. Inclusion in new plants is more simple. Capital costs for equipment are relatively low and and usually can be traded off in excess capacity for peak loading. Operating costs are higher for chemicals and controls.

3 Upgraded effluent quality necessarily costs more. There are no free-rides. Going from 80 to 95% phosphorus (P) removal increases cost by 50% or more. The increased cost provides more complete control of general and specific cleanup with improved contingency control as an added benefit. Major upgrading costs are for applied chemicals once equipment and control are available.

4 The owner must make the choice of routes to meet requirements. Reduction in solids and BOD may be pivotal. Success demands the owners commitment to:

a Provide capable operation 24 hours/ day, 365 days/ year

b Provide capable laboratory support for operational control not just "for the record"

c Provide a budget of about 5¢/ 1, 000 gallon of treated flow.

5 The degree of treatment dilemma forces consideration of functional versatility and frequent upgrading. Coagulation is one means of providing a high degree of upgrading capabilities. Use what you need today and adjust for tomorrow's needs.

II COAGULANT APPLICATION SITES

A Many possibilities exist. Choice depends upon facility, equipment, layout, and requirements to be met. Ingenuity is required to pick the best location for a given situation on a cost-effective basis.

B Coagulant addition during primary clarification provides the highest yield of solids separation per unit dosage at a reasonable percentage removal.

1 Increased oxygen demand removal from about 30 to 50% may be a clear choice in an overloaded facility.

2 Primary sludge is easier to handle and much more concentrated than equivalent solids as a secondary sludge.

3 Remaining turbidity in the primary effluent is more readily cleared in subsequent units.

4 The raw influent has the lowest fraction of ortho/total P of any liquid in process. Phosphorus or solids removal may not be as complete as at later stages of operation without high coagulant dosage.

C Coagulant Addition in a Biological Unit is Attractive

1 Coagulation during trickling filtration may favor blotchy appearance and increased sloughing. Plugging does not appear to be a problem. Mixing characteristics and other factors decrease the cost/benefit ratio of coagulation in the trickling filter.

2 The mixing, detention, and floc characteristics of activated sludge aerators favor coagulation benefits.

a Coagulants may be added at the inlet, middle or outlet zone of aeration or en route to the final clarifier. The last is most popular.

b Tenny and Stumm(6) favored coagulation during contact stabilization.

c Effect of metals on return sludge and biota is unclear(7). The floc characteristics may favor decreased chemical usage but may also impede contact of substrate and critters, oxygen transfer, and off gas release. The control of pH in the cell environment during coagulation is imperative to avoid biological problems. This generally means in place measurement in a dynamic situation where CO_2 equilibria is involved.

d Coagulation of activated sludge tends to increase sludge concentration, provide greater solids capture, speed up sedimentation and decrease phosphorus recycle. Coagulation is the most direct and feasible correction for "bulking."

D Coagulation in the final clarifier is a last chance situation, but, quite effective.

1 Prior stabilization decreases surfactants and general resistance to liquid-solids separation.

2 Phosphorus generally exists mainly in the ortho form that can be precipitated.

3 If the underflow is returned to the primary clarifier it favors liquid-solids separation in the primary.

4 Producing a clear well-stabilized overflow is obvious proof of good operation. Correction of earlier poor guesses is good public relations too.

E Multi-point application of coagulants makes it possible to remove most of the loading at the most favorable stage and to polish the discharge at its most favorable stage with lower chemical requirements and improved control.

F Emphasis on upgrading existing facilities precludes consideration of filtration and other advanced treatment in this outline.

III TRIAL EFFORTS--HOW MUCH? HOW LONG?

A The vagaries of wastewaters, situations and conditions make it hazardous to generalize the response of any particular "improvement." Joe may swear by it--Tom, at it. Try it--preferably on small scale under conditions anticipated for use. Become familiar with likely events in response to your particular situation and its changes. If the results include a horrible mess--hopefully--it's a small one.

B Jar tests are an ever present help in time of trouble--a vital but treacherous ally. Don't just "follow the book"--THINK.

1 The "stare and compare" technique between jar test and full scale operations is invaluable. Don't expect jar test information to be transferable to full-scale if the situations are incomparable.

2 Coagulation occurs within seconds of contact. If it isn't mixed, it hasn't contacted, and the coagulant addition benefit is reduced. An auxiliary flash mix is highly desirable to promote

desired reactions including substrate rather than interactions of coagulant and water.

3 Flocculation or gathering together of small particulates takes place under lower energy input situations. Time and energy are important--you can favor agglomeration or break it up.

4 Settling is a dynamic situation. Some movement is beneficial to pick up stragglers, to much produces stragglers.

5 Practice, practice, practice.

C Other laboratory tests are described(8, 9) Unfortunately information in the book isn't much good until it's applied.

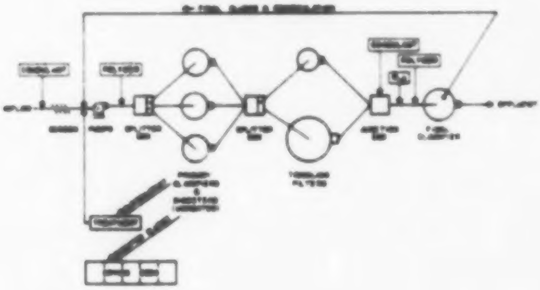
D Full plant (or an isolated module) trials generate confidence in results. If carefully planned and executed in line with pre-tests and other information, the trial is not likely to bankrupt the organization or create an intolerable mess. Figure 1 describes the full scale test at Richardson, Texas(10)

IV CHEMICALS: WHICH ONES? WHAT CAN THEY DO? HOW?

A Coagulation for use in conventional treatment usually means an iron or aluminum salt. Lime is an excellent coagulant but it's generally more valuable in physical chemical treatment rather than for supplementation of conventional treatment. Lime may be required along with iron or alum salts to maintain the biological systems in a favorable pH range for growth.

1 Iron salts as the chloride or sulfate are more effective in the oxidized form (Fe^{+3}). Waste acid from scale removal containing reduced iron may be neutralized and aerated to provide an effective coagulant.

2 Aluminum salts may be applied as alum or sodium aluminate. Alum or iron addition tends to reduce pH upon hydrolysis. Sodium aluminate is more expensive but it includes alkalinity and tends to raise mixture pH. Its use may relieve neutralization cost in low alkalinity water.



	Diam (Ft)	Depth (Ft)	Circum (Ft)	Area (Sq Ft)	Volume (Cu Ft)	(Gal)
Primary Clarifier No. 1	40	8	126	1257	10,056	75,200
2	40	10	126	1257	12,570	94,000
3	40	10	126	1257	12,570	94,000
All Primary Clarifiers	--	--	378	3771	35,196	263,200
Final Clarifier	70	6	220	3848	23,088	173,000
Filter No. 1	84	6.5	---	5542(1)	36,000	---
2	120	6.5	--	11310(1)	73,500	---
Filters Combined	---	---	---	16852(1)	109,500	---
Digester No. 1	40	14.3(2)	---	1257	13,000	135,000
2	40	14.3(2)	---	1257	13,000	135,000
3	40	14.3(2)	---	1257	13,000	135,000
Digesters Combined	--	---	---	---	39,000	404,000

Sludge Drying Beds 12,000 Square Feet

(1) Area in acres: 0.127, 0.260 and 0.387, respectively
(2) 14.3 Effective, 18.0 SWD, 15.8 Clear @ Center

Figure 1. Treatment Plant with Chemical Precipitation Facilities(10)

3 Coagulant aids, as the name implies, may be used to favor agglomeration of destabilized particles or for the same purpose after coagulation. Many forms are available and many more are under development for specific needs.

B Available Forms of Coagulants

1 Liquid solutions of iron or alum are easiest to use. Solution preparation is a tedious and time-consuming job. Cost,effectiveness, flexibility, and ease of handling all point toward use of the solution form. Transportation cost of added water for long distances may rule them out.

2 Polymers generally are shipped in dry form.

C Technical Details on Coagulants

1 Alum as a 48.5% solution is 4.37% Al and weighs 11.1 lbs/gal. See the suppliers data for other preparations.

2 Ferric chloride is available in 35% to 45% solutions. (The 45% solution freezes at 45°F and 37% solution freezes at 15°F.) The 40% solution weighs 11.9 lbs/gal and contains 16.4% Fe.

D Functions

1 Calcium iron and aluminum salts hydrolyze to form insoluble precipitates with soluble ortho phosphate and poly phosphates(12).

2 Metal salts act upon stable suspensoids to destabilize the forces maintaining dispersion such as the reduction of the surface charge on hydrophobic dispersions and dehydration of the water layer on hydrophillic dispersed components.(11, 13, 14, 15)

E Coagulation Reactions and Kinetics

1 Hydrolysis and reactions of metals in coagulation are very rapid--probably less than one second.

2 Metal radicals are complicated(16).

3 Effects of biologically produced polymers are obscure(17, 18).

F Flocculation Reactions and Kinetics

1 Flocculation primarily is a gathering together of fine particles produced by precipitation or coagulation into aggregates large enough to separate effectively.

2 Flocculation is a lower energy situation and time required is a matter of 5 to 20 minutes.

3 Coagulant aids, natural polymeric materials and coagulants function to glue particles together, to provide a skeleton for a dense floc and to "sweep up" straggler particles. Particle contact is essential at an energy level favoring agglomeration rather than dispersion.

G Inferred Physical Arrangement

1 Intense flash mix for 2-30 seconds while adding coagulants.

2 High energy flocculation for 1-5 minutes (add polyelectrolytes near midpoint).

3 Low energy flocculation - 5 to 20 minutes.

V HARDWARE: TYPE, SIZE, USAGE

A Coagulant Feeding Equipment

1 Many types of variable discharge pumps are available. Positive displacement pumps are effective. Diaphragm, plunger, gear, progressing cavity types are alternate choices.

2 Use pumps designed for 500 psi service, install a 40 psi back pressure valve and neglect head loss in following piping.

3 Backpressure helps seat check values. Double check values are good.

4 Pump sizes that are too large makes low range control difficult. Some pumps have interchangeable heads for selection of suitable control ranges.

5 Ratio control is helpful in addition to percentage output control.

6 Proper selection of pump materials permits interchange of coagulants and polymers.

7 Control units

a Pumpage record from operators log is one method of manual control.

b A magnetic flow meter from storage, gravity flow, a temperature indicator recorder (TIR) and a set point controller on a valve ahead of the feed point is indicated in Figure 2.

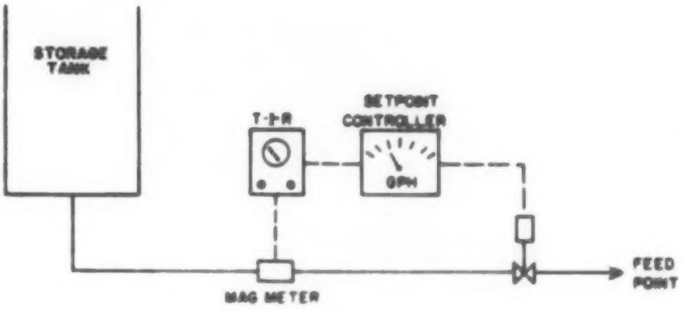


Figure 2. Gravity Chemical Feed System

c A compound loop control is indicated in Figure 2 including a flow meter, phosphorus analyzer and an integrator and set point selector and chemical pump controller.

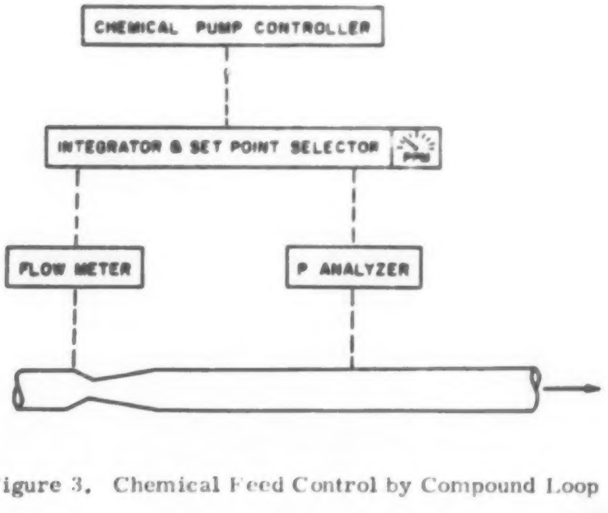


Figure 3. Chemical Feed Control by Compound Loop

d Automatic controls permit closer trimming of dosage and flow demand but require close control of sensors, set-points, instrumentation and mechanics. It takes more know-how to keep them automating.

B Flash Mixing

1 This is a most grevious shortcoming of many chemical operations(20, 21).

2 Contact of coagulant, water, and substrate is necessary before the reaction is complete or coagulant hydrolysis products may not include substrate.

3 Hydraulic jump boxes, drop box or manhole, pump intake or baffled basins provide mixing but not flash mixing.

4 A propeller or turbine mixer, jet, vortex or other devices(22, 23, 24) with an in-line approach providing an input G of 800-1000 for up to 30 seconds is important.

5 For an electrically driven mixer:

$$G = \left[\frac{(WHP)(550)}{\mu\ V}\right]^{0.5}$$

where: (WHP) = delivered water HP

or $= \dfrac{(KVA)\binom{\text{Motor}}{\text{Efficiency}}\binom{\text{Power}}{\text{Factor}}}{0.746}$

μ = absolute viscosity, (0.2×10^{-4}) at 70°F

V = mixed volume

6 For a baffled basin or other head loss unit where

H = 1 ft. head loss

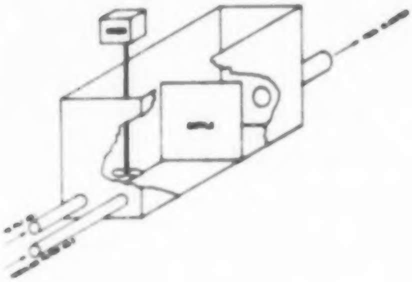
T = 31.2 sec. detention time

$$\text{then: } G = \left[\frac{(62.4)(1.0)}{(31.2)(0.2 \times 10^{-4})}\right]^{0.5} = 316^{-1} \text{ sec}$$

This is much less than the recommended 800 to 1000 G. The chemical reaction may be complete in 5% of the time allowed.

a To produce a G of 1000 the preceding drop box time would be 3.12 sec and box volume would be 4.5 cu. ft./MGD--an interesting design problem.

b An effective device is outlined in Figure 4. This is an open flash-mix box. Mix at the inlet end leaving at least three fold volume for future needs. A junction box may serve if an appropriate mixer is provided.



Flow Range		Coagulation	Flocculation Time (Minutes)		
MGD	GPM	(Minutes)	High Energy	Low Energy	Total
1	700	1.42	3.71	20.6	24.3
1.5	1,050	0.95	3.01	19.1	22.1
2	1,400	0.71	2.80	14.3	17.1
2.5	1,750	0.57	2.30	11.4	13.7
3	2,100	0.48	1.91	9.5	11.4

Figure 4. Junction Box Modified to Flash Mix[10]

c Proper mixing in an aeration device depends on its mixing characteristics. Add coagulants at the point of maximum energy and hope.

C Flocculation

1 Water plant technology(25, 26, 27) serves for wastewater use.

2 High energy flocculation occurs at the outlet end of the mix box, piping, or clarifier inlets. Clarifier inlet hydraulics are critical.

3 Low energy flocculation occurs in the sludge blanket.

4 Baffles, paddles, etc., may not be needed.

5 An activated sludge aeration tank may be a good flocculator.

D Piping

1 Polyvinyl chloride (PVC) and Fiberglas Reinforced Plastic (FRP) protected from freezing is required for coagulants.

2 Provide flushing tees following pumps.

3 Accumulators are not necessary; manual air blow-offs recommended. These are also useful as sampling and calibrating ports.

4 Use strainers on pump suctions and large lines.

5 Install a dilution line in the polyelectrolyte pump discharge for about 20 gpm water input. A 50 ft. run of pipe, a jet, or turbulence section in a shorter pipe will provide mixing.

6 Include hose stations for spillage washdown.

E Storage Tanks

1 Filament wound or layup fiberglas tanks from a good supplier are dependable. Natural amber or colored choices are available.

2 Size for a 7-10 day supply on a 2/1 mole ratio. About 3,000 gal/MGD for alum (equals three weeks for $FeCl_3$ or aluminate) a 6,000 gallon tank will accept 5,000 gallon tank truck lots.

3 Use a polyurethane foam pad between mounting slab and tank unless weather dictates a heated pad to keep coagulant warm.

4 Install a sight gauge, manhole, fill inlet with snap coupling, pump suction, vent, flush-bottom drain.

5 Polymer tanks usually sized for a 2-3 day supply of 0.5% solution. This would be 400 gal./MGD at a 2 mg/1 dose. Can and probably will want to dilute below 0.5% before introduction.

6 Use same tank fittings as for coagulant (E4) plus an over powered mixer for 1,000 cp viscosity dissolving requirements. Water connectors for fill inlets and dispenser funnel required.

7 Consider shelter for polymer tanks--the operator will appreciate it.

8 Provide an emergency reserve tank for coagulant and polymer--a 4 to 6 hour supply.

VI Dosage selection control is the key to success; this controls both cost and performance.

A In General

1 Inject coagulants at full strength (to prevent premature hydrolysis) into an intense dispersion zone. Use one inlet pipe per mixing unit up to 8 to 10 MGD of treated flow.

2 Inject polymers in a diluted stream with thorough dispersion at a moderate energy level. Have several available inlet ports. Use a multiport header for treated flows above 3 MGD.

B The key parameters in dosage control is the mole ratio of coagulant requirements versus material to be removed on a reacting equivalent basis.

1 The coagulant expressed as metallic aluminum (Al) or iron (Fe) as a pound molar form makes it possible to change coagulants or coagulant mixtures and maintain effective dosage by correcting for the particular mixture on the basis of its metallic percentage or decimal fraction. One pound mole for Al = 27 pounds; Fe = 56 pounds.

2 For purposes of this outline, phosphorus (P) is used as the index of material to be removed. One pound mole of Al or Fe is equivalent to 1 pound mole of P. The molar weight of P is 31 or 31 pounds = 1 pound mole.

3 If the treated flow contained reactable phosphorus only, it would be possible to remove P predictably by adding one pound mole of Al solution for each pound mole of P. Municipal wastewaters are complex mixtures of different components; incomplete or competing reactions always require more coagulant than is indicated by any one measured index such as P. Experience shows an effective ratio to be 1.5/1.0 to 2.0/1.0 in coagulant (C) metal to P load (a C/P mole ratio).

4 For example:

For each 1 MGD of flow containing 10 mg P/liter (equivalent to 10 lbs. P/M pounds of flow).

a Calculate P load in terms of pound moles/day

$$\frac{10\text{ lbs. (P)}}{\text{M lbs.}} \times \frac{1\text{ MG}}{\text{D}} \times \frac{8.34\text{ lbs.}}{\text{G}} \times \frac{1}{31}\,\frac{\text{mole}}{\text{lbs.}}$$

Cancelling common terms and computation

$$= \frac{83.4}{31} = \frac{2.69\text{ or }2.7\text{ lb. moles of P}}{\text{D}}$$

b Knowing the phosphorus per MGD entering the plant in pound moles provides an entry for coagulant dosage, i.e., for a 2.0/1.0 C/P ratio to produce a satisfactory effluent--use it.

2 × 2.7 lb. moles = 5.4 lb. moles in terms of required metal dosage/MGD.

c This metal may be added as a dry salt or as a water solution. Suppose we decide to add it as an alum solution containing 48.5% alum, and 4.37% Al weighing 11.1 lbs/gal. Dosage becomes:

$$5.4\text{ lb. mole Al (C)}\frac{27\text{ lbs.}}{\text{lb. mole}} \times \frac{\overset{\text{Alum}}{100\%\text{ soln.}}}{4.37\%\text{ Al}} \times \frac{1\text{ gal.}}{11.1\text{lb.}}$$

= 302 gal. alum solution/MGD coagulant dose rate.

d The same number of lb. moles of iron coagulant could be used but the volume would vary with the iron content of the applied coagulant. For a 40% $FeCl_3$ solution containing 16.4% Fe and weighing 11.9 lbs/gal:

$$5.4\text{ lb. moles Fe(C)} \times \frac{56\text{ lbs.}}{\text{lb. mole}} \times \frac{\overset{\text{soln.}}{100\%FeCl_3}}{16.4\%\text{ Fe}} \times \frac{1\text{ gal.}}{11.9\text{ lbs.}}$$

= 155 gal. $FeCl_3$ soln/MGD coagulant dose rate.

Hourly dose rate = 155 × $\frac{1}{24}$ or 6.5 gal/hr. for the average dose rate of 40% $FeCl_3$ solution.

5 The data in Item 4 were based on average flow and concentration basis. The average doesn't hold for either flow or concentration on municipal wastewaters.

a Figure 5 shows the distribution of influent P by the hour. Relationship of soluble P and total P are indicated in the same graphic. This means coagulant dosage adjustment according to the time of day to prevent overdosage at night and underdosage during the peak loading.

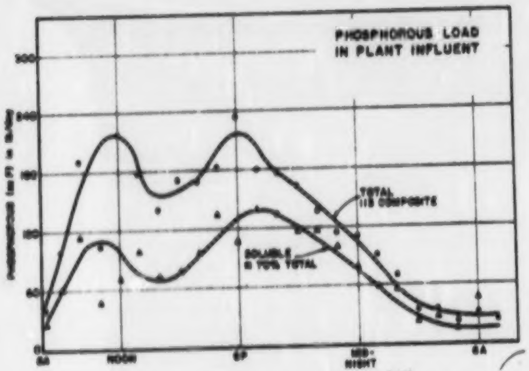


Figure 5. Phosphorous Income Plot(10)

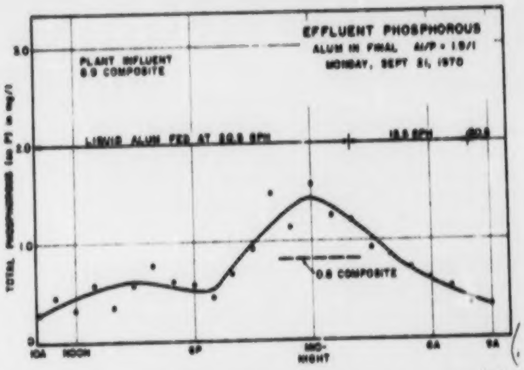


Figure 6. Too Few Chemical Pump Settings Give Poor Control(10)

b Adjusting coagulant dose rate 3 to 5 times/day is more likely to match coagulant dose and requirements.

c Measurement of soluble P requires coagulant dosage adjustment to meet total P requirements. In Figure 5 the soluble P was about 70% of total P; therefore, a sensor detecting soluble P only requires a coagulant dosage correction of 100/70.

d Daily income of P also may vary. Experience(10) showed high P on Saturday, low P on Sunday and fairly regular loading on weekdays. Check your loading pattern.

e Automatic control relieves the operator from manual adjustment. The operator is obligated to adjust setpoints for control in line with effluent quality and to validate control operation.

6 Effluent P is a direct result of coagulant dosage control. Figure 6 shows the effluent P concentration by the hour. A twice daily dosage adjustment was insufficient to maintain effluent P below 1 mg P/liter.

7 Dosage corrections must be varied until the significant peaks are eliminated. Figure 7 indicates dosage changes and effluent P response. Stay on a given schedule at least five days--the first day may not show adjustment effects meaningfully. Don't overcompensate. Figure 8 shows effects of split dosage to raw and final clarifiers. The dosage is the same but better control is evident.

8 Figure 9 shows a series of results indicating effluent P versus mole ratio of C/P. These results are characteristic of the influent, the operation and the situation. Some change may be expected in other situations.

C Polymer coagulant aids are intended to improve solids capture and may reduce coagulant requirements. If total cost is reduced enough to pay for the coagulant--use them.

1 Acquire information on recommended available polymers.

2 Try several by jar test with your influent and operations.

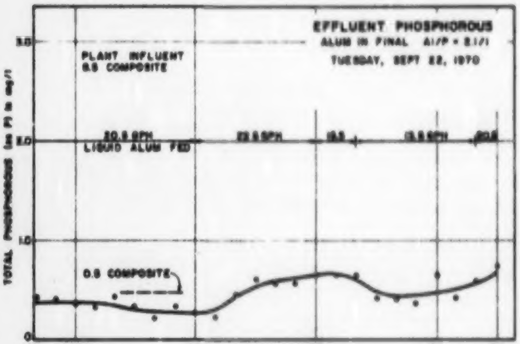


Figure 7. Effluent Phosphorous (Tuesday, 9/22/70)

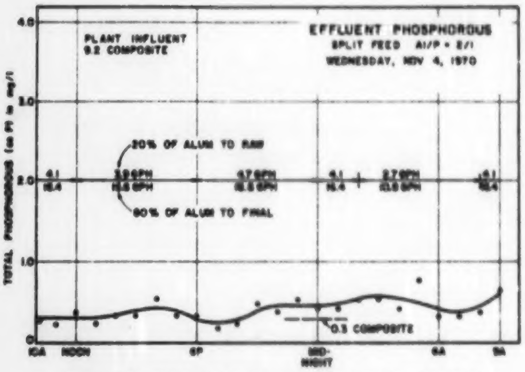


Figure 8. Effluent Phosphorous (Wednesday, 11/4/70)

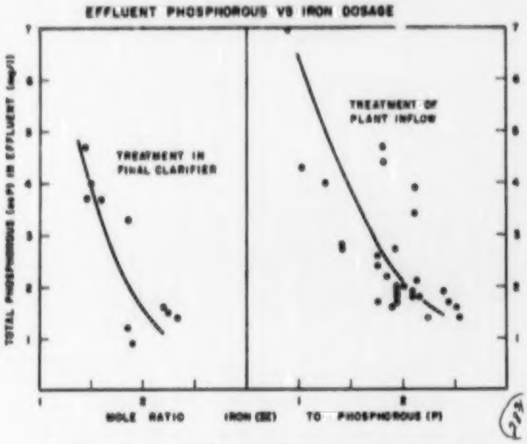


Figure 9. Study of Results Allows Selection of Desired Mole Ratio

3 Try the more promising items in your plant.

4 A typical dosage is 1 mg/liter or less. The proper dose and dispersion is indicated by floc behavior, "Quicky" tests like the "slick fingers," effluent turbidity, and floc appearance are useful.

5 Feed rate usually is constant, possibly, reduced at night.

VII SAMPLING AND ANALYSIS

A Coagulation during plant operations requires more laboratory control than the same plant without coagulation. Whatever has been done will continue and extras are inevitable. As in all upgrading operations the tests for the record may not be the same as those for control but both are essential. A good look at what you are doing and why may lead to unexpected improvements.

B Figure 10 lists possible tests for conventional treatment and additional tests for phosphorus removal according to stage of treatment or sample.

ANALYSES FOR CONVENTIONAL TREATMENT

	RAW	PRIM EFF	FILT EFF	FINAL EFF	RECIRC	SLUDGE	SUPERNAT
FLOW	●				●	●	●
TOT SOL	●	●	●	●	●	●	●
TOT VOL SOL	●	●	●	●	●	●	●
SUS SOL	●	●	●	●	●		●
SUS VOL SOL	●	●	●	●	●		●
SET SOL	●	●	●	●			
BOD	●	●		●			●
DO	●			●			
COD	●	●	●	●	●		●
PH	●	●	●	●		●	●
TEMP	●	●	●	●		●	●

ADDITIONAL ANALYSES FOR CHEMICAL TREATMENT

	RAW	PRIM EFF	FILT EFF	FINAL EFF	RECIRC	SLUDGE	SUPERNAT
PHOS	●	●		●	●	●	●
ALK	●	●		●		●	●
FE	●	●		●	●	●	●
AL	●	●		●	●	●	●
SO4	●	●	●	●		●	●
CL	●	●		●			
TURB	●	●		●			

Figure 10. Chemical Precipitation Involved Added Analyses

1 Some of these are more important than others. Control tests may be spot checks on catch samples by the most rapid available routine. Record tests may be based upon composite proportional samples by a referee procedure. Good control means more tests and more meaningful testing with rapid feedback to process control.

2 Coagulant analysis means analysis of both cation and anion concentration. Alkalinity may not be necessary if pH problems are unlikely. When alkalinity decreases to less than 50 mg/liter watch it carefully and readjust promptly.

3 Effluent turbidity is a rapid and handy test for dosage control. Sludge blanket level procides an early warning of impending problems.

VIII SLUDGE HARVESTING AND DISPOSAL

A Clarifiers: Make it drop--and stay down.

1 Nine feet SWD for Trickling Filters, 12 feet for activated sludge. 500 gpd/sq. ft. surface overflow rate avg. floc, <1,000 peak flow, good inlet system.

2 Preserve blanket, keep recirculation low in trickling filters. Watch operations particularly floc behavior.

B Sludge Treatment and Disposal--Perennial Headache, Doubled

1 Sludge digests well, percent solids higher than conventional.

2 Be courageous with digesters; provide heat and mix. Consider thickening, but treat overflow carefully--colloidal P strays.

3 The volatile solids may increase 50 to 100% by weight--and carries added minerals with it.

4 Fe sludge on drying beds better than Al > conventional. Drying time 1/2. Don't add too much at one time, maintain >6" of sand for a good underflow.

5 Chemical cost reduced for vacuum filter or centrifuge operation step.

C Supernatant, Rogue Recycle, Other Happy Thoughts

1 Supernatant treatment before recycle:

a Fill and draw tanks, or flow through

b 250 mg alum/liter plus 20 min. air then settle

c Sludge to dewatering or digester; return clear water. It works.

d Consider aeration, 200-400 mg lime/liter, aerate settle. Can strip NH_3 at high pH.

2 Iron leakage: (Alum does too but it isn't red.)

a Worst on final settler treatment--use split does.

b Polymer reduces escaping colloids, Figure 11.

c Use most of the iron in the primary. Give it time to be gathered. Aeration time helps.

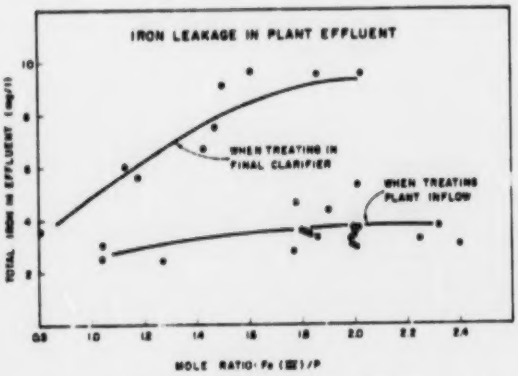


Figure 11. Iron Leakage in Plant Effluent

3 Other radicals have an impact too:

a 1 lb. Al adds 5.35 lb. sulfate

b 1 lb. Al, as aluminate, adds 0.85 lb. sodium.

c 1 lb. Fe^{+2}, as chloride, adds 1.26 lb. of chloride.

d 1 lb. Fe^{+3}, as chloride, adds 1.91 lb. chlorides.

e 1 lb. Fe^{+3}, as sulfate, adds 2.58 lbs. sulfate.

IX COSTS

A Capital Investment

1 FRP Tanks: $1/gal. up to 1,000 gal.; $0.60 above.

2 Pipe, fittings, valves: $1,000/MGD, don't scrimp here.

3 3 HP Mixer, starter $1,000.

4 Chemical feeders, starters and auto. capability $1,500.

5 Concrete, baffles, samplers, laboratory equipment, controls, electric gear, shelter: depends on the situation.

6 Capital costs are a small part of the total.

B Chemical Costs: This is where the money goes.

1 Liquid alum 48.5% solution, at $0.24/lb. Al plus freight ($0.12/lb/250 mi.).

2 Sodium Aluminate, near $0.35/lb. Al.

3 Ferric chloride $0.12/lb. plus freight (about $0.05/lb./250 mi.).

4 Spent pickle liquor: no guesses.

5 Polymer costs vary widely. About 1¢/1,000 gal. is a typical upper limit.

6 Your cost estimate: (Pincome)×(C/P ratio) × (unit cost).

C Other Costs

1 Power nominal about $200/yr./MGD.

2 Man power: With intelligent 24 hour per day operation the same personnel numbers can be trained to handle coagulation. If 24 hour operation wasn't available it should be. For automatic control a good instrument man is needed or on call.

3 Laboratory Support: If lab staff is available possibly one more. Usually a good lab man can handle both conventional and coagulation tests with technician help by operating staff.

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This outline was prepared by James E. Laughlin, P.E., Shimek, Roming, Jacobs and Finklea Construction Engineering, 2118 Adolphus Tower, Dallas, TX 75202.

ALTERNATE METHODS FOR NITROGEN AND PHOSPHORUS REMOVAL

I INTRODUCTION

A Technology of nitrogen (N) and phosphorus (P) removal from wastewaters has materially improved in the past 5 years. Many process possibilities are available.

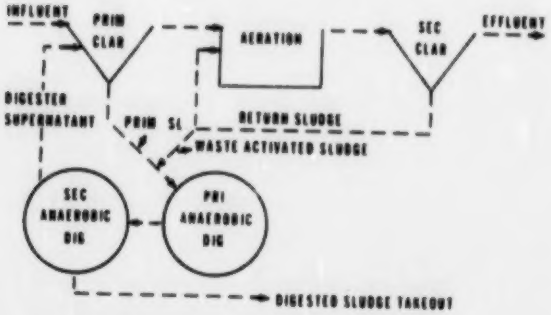
B Some of these have been demonstrated in full scale plants, some in large pilot scale units and others only at bench scale.

C Processes favoring nitrogen and phosphorus removal may be based upon biological chemical or physical factors or combinations thereof. This discussion briefly considers several possible removal procedures according to the dominant discipline involved.

II BIOLOGICAL METHODS OF N & P REMOVAL

A Any conventional treatment system will remove some of the N and P. The net removal is relatable mainly to the amount of sludge takeout, characteristics of loading, oxygen excess, possibilities for precipitation, growth and feedback are important.

FIG. 1



CONVENTIONAL WASTEWATER TREATMENT

FIGURE 1

(BARTH HAP REM)

B Phosphorus Removal

1 All plants reporting efficient removal of P do not recycle supernatant as in Fig. 1.

2 Digested solids takeout is the only P removal route in usual operations.

3 Several theories are used to account for P behavior during treatment.

   a Rapid synthesis and luxury uptake-San Antonio

   b Chemical precipitation-Jenkins

   c Combination of a & b-Baltimore

   d Leaching in process confuses a, b, c-Levin

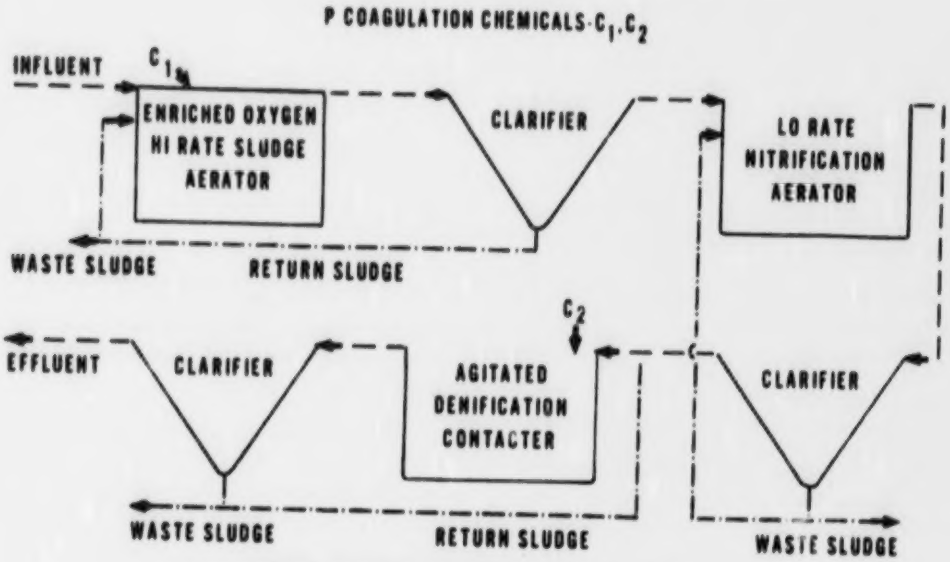
4 Controlling factors in P removal

   a Suitable C/P ratios
   b Suitable F/M ratios
   c Rate of air supply and process DO
   d Rapid sludge return
   e High MLSS in process
   f Either minimize leaching or take out leachate.

5 Technology for biological removal of P has not advanced to the point where high removals are possible in all situations.

C Nitrogen removal

1 Nitrification and denitrification in suspended growth or attached growth systems is the most feasible route available on a cost/benefit basis for nitrogen removal.


P COAGULATION CHEMICALS-C_1,C_2
INFLUENT
C_1
ENRICHED OXYGEN HI RATE SLUDGE AERATOR
CLARIFIER
LO RATE NITRIFICATION AERATOR
WASTE SLUDGE
RETURN SLUDGE
C_2
EFFLUENT
CLARIFIER
AGITATED DENIFICATION CONTACTER
CLARIFIER
WASTE SLUDGE
RETURN SLUDGE
WASTE SLUDGE

THREE SLUDGE SYSTEM: N&P REMOVAL

FIGURE 2

(BARTH N&P)

2 The atmosphere (80% molecular nitrogen) acts as a nitrogen sink from which combined nitrogen is produced and returned. Nitrogen gas is converted to combined nitrogen by biological chemical and physical actions, such as electrical discharge, biological action and commercial nitrogen fixation. Fixed nitrogen recycles back to the atmospheric sink largely by biological action (1).

3 Two major considerations are essential:

a The conversion of organic and ammonia nitrogen (unoxidized N) to nitrite or nitrate (oxidized N) must be as complete as possible for effective removal.

b Conversion of oxidized nitrogen to moledular N_2 (denitrification) requires the presence of an organic substrate to act as an hydrogen donor to force the dissolved molecular oxygen down and cause the nitrate oxygen to be used for metabolic purposes with release of nitrogen gas(2). Methanol is the most common organic substrate used in engineered systems because it is cheap, effective and does not produce excessive growth (3).

4 Many natural and engineered systems have been effective for some degree of nitrogen removal on an incidental or deliberate basis.

a Stream removal of N has been observed for years. Surface water generally stabilizes with time to produce the organisms and oxygen for nitrification. Contact with benthic deposits provides denitrification.

b Facultative ponds , lakes and reservoirs undergo similar sequences. Vertical circulation improves nitrogen removal in stratified waters.

c Fine media filtration encourages nitrification in the upper parts of the bed with denitrification in the lower strata of filter media. Pressurized filters up to 70 psi with flow rates of 7 gal/sq ft/min have been used.

d Large media filtration, such as trickling filters, tend to produce nitrification in surface slimes, denitrification near anaerobic zones. Deep beds encourage the process. Up flow operation for denitrification is favored.

e Suspended growth reactors achieve nitrification and denitrification to some extent providing the system has a periodic cycling of low and high loading and the sludge age permits nitrification (>5 days).

f The most efficient removal of N is achieved in a 3 sludge system where most of the organic load is removed in a high rate sludge, nitrification is achieved in a low rate system of long sludge age, and denitrification is accomplished in an agitated sludge system without aeration and with the addition of methanol as an hydrogen donor for nitrate oxygen. Chemical addition of iron or aluminum may be used in stage addition to obtain P removal in the same system. The chemical addition also improves sludge characteristics and greatly retards Precycle (Figure 2).

III CHEMICAL METHODS: N & P REMOVAL

A Many methods are available that may be used. Upgrading effluent quality with respect to solids, oxygen demanding materials, N, P and other possible residuals requires specific knowledge of the chemistry, the cost/benefit ratios and control technology.

B Phosphorus Removal.

1 Removal of both nitrogen and phosphorus may be achieved as a side stream process on digester supernatant by formation of magnesium ammonium phosphate ($MgNH_4PO_4$)

a Addition of magnesium chemicals and alkali results in $Mg(OH_2)$ formation in the digester supernatant recycle stream or sludge filtrate. The insoluble $Mg(NH_4)PO_4$ settles out and can be removed by solids liquid separation to form a low grade fertilizer and an upgraded recycle stream.

b Side stream treatment relieves a major fraction of recycled load in secondary treatment facilities.

2 Ion exchange is considered by Dobbs in another outline.

a The phosphate anion may be removed by anion exchange resins. Selectivity of the exchange resins presently means that sulfates nitrates chlorides are significant interferences with P removal resins.

3 Lime coagulation of P is considered in other outlines and is currently favored because of its effectiveness in removal of P, and other solids; it is available, cheap, recoverable, and reusable by conventional techniques.

a This process is currently favored for physical chemical treatment in conjunction with carbon adsorption in several large scale plant designs.

b The process is controllable, efficient, and feasible but requires large sludge handling facilities, lime recovery to reduce solids disposal problems, and is not effective for nitrogen removal. Sulfide production in the following carbon bed requires special consideration.

4 Lime treatment in the primary clarifier followed by biological treatment is used as the Dorr-Oliver PEP process. This is attractive for upgrading overloaded conventional systems by coagulation of most of the solids and P in the primary with polishing treatment in a conventional secondary system. Control pf PH, and usually recarbonation, is essential to protect the biological system. Nitrogen control is possible with this combination.

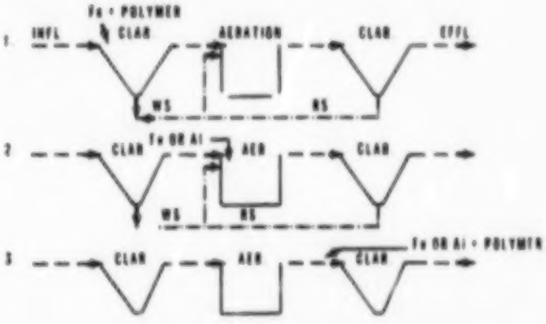
Lime coagulation of the effluent following conventional treatment has been used for P removal (considered in other outlines).

a Lime requirement for P removal is a function of influent alkalinity.

b Nitrification of the effluent during treatment reduces lime requirements by about 7 wt. units of alkalinity per unit NH_3-N converted.

6 Aluminum and Iron salts are effective coagulants for P. These have been used in large scale applications for water treatment for years. Many plant scale operations on wastewater are underway. Lanthanum precipitation of P favors lower residuals but currently has been limited to laboratory investigation. Recovery of iron and aluminum from sludges for re-use currently is not attractive. Major advantages of iron and aluminum salts for P removal are the possibilities for P removal in conventional plants, supplementation in existing facilities, backup or contingency use, or small plant application where chemical recovery is not indicated. Polymer application with coagulants often improves behavior.

a The Dow process employing Iron or Aluminum coagulants plus polymers may employ any one of 3 treatment schemes: (Figure 3).



DOW PROCESS FOR PAND SOLIDS REMOVAL
FIGURE 3

(BARTH)

(Activated Study)

1 Scheme 1 was used at Grayling MI, 2 was used at Benton Harbor MI and 3 was used at Lake Odessa TX (plant scale) among others (Trickling Filtration).

2 Polymers are not needed in Scheme 2 because the surface area of the biological floc and time of contact favor good coagulation with economical use of chemicals. The ratio of metal to P and probable removal of P is likely to be:

1/1	80 to 85% P removal
1/2 to 1.5/1	85 to 90% P removal

3 Greater than 90% removal of P usually requires some form of effluent filtration since P removal is a solids removal process. The separated dispersed solids in the effluent from most treated effluents are likely to contain from 1 to 3% P dry basis.

a Split coagulant dosage at two of the three chemical addition points indicated or polishing treatment in following operations commonly provide lower effluent P with more efficient use of chemicals.

b Split dosage in the primary and denitrification stage was advantageous at Manassas, VA.

C Chemical N removal processes.

1 The addition of ferrous iron plus a catalyst to nitrates results in the production of $N_2 + Fe^{+3}$.

a Current status calls for high concentrations of Fe and catalyst to make the process work.

b Extraneous salts complicate control due to formation of iron-anion complexes.

c Progress is being made.

2 Ion exchange of both NH_3-N with cation exchange resins and NO_3-N with anion exchange resins is considered in other outlines.

a Nitrogen removal by ion exchange commonly favors weak base or weak acid type resins.

b Resin selectivity for nitrogen removal is better than for P removal.

c Ion exchange for N requires a highly clarified effluent to prevent blinding. Regenerant disposal, removal of competing ions, capital and operating cost are major problems.

d Prospects are improving with exchange material development.

3 Chlorination of Ammonia in wastewaters has long been recognized with respect to effects upon chlorine requirements for disinfection. Complete reaction of chlorine and ammonia result in production of nitrogen gas and hydrochloric acid.

a It requires a wt ratio of about 10:1 of chlorine/NH_3-N.

b The reactions of NH_3 and Cl_2 are predictable and controllable.

c Presence of organic N and other chlorine reacting material raises chlorine requirements and complicates N removal efficiency.

d The need for neutralization generally raises effluent total solids materially.

e The possible residual toxicity of side reaction products or reactants of chlorination are receiving increasing attention. Effects are difficult to evaluate and to control.

IV PHYSICAL PROCESSES: N & P REMOVAL

A Phosphorus removal

1 Solids- liquid separation of precipitated P by gravity separation, screening, filtration, centrifugation, etc., are the terminal step of other processes for P removal.

2 Distillation is a feasible route for separating most forms of combined P from the aqueous phase. The P remains in the unvaporized fraction.

3 Electrodialysis has been included in other discussions as a practical means of separating an aqueous solution containing dissolved salts into a concentrated phase and dilute phase of the salts included. Both N & P salts are subject to this distribution.

a The distribution does not provide complete separation under feasible conditions; hence, is more effective for reduction of an unacceptable concentration to an acceptable concentration.

b The disposal of the reject concentrated brine is a major problem.

4 Reverse osmosis has been developed to obtain about 90% separation of both N and P from wastewaters. It requires moderately high pressures, large membrane areas and a high degree of influent clarity to be effective.

5 Electrolysis of water has been used in Sweden to remove P.

a High solute concentrations are required to make this process economically attractive.

b The P may be coagulated as $Mg(NH_4)PO_4$ with the formation of $NaOH$ and Cl_2 in process. The effluent requires neutralization for pH control but is likely to be well disinfected.

c Sacrificial electrodes of Al or Fe result in production of coagulating chemicals in process.

B Nitrogen Removal:

Many of the physical processes used for phosphorus removal are also effective for N removal. They may differ with respect to the form or mode of removal.

1 Solids-liquid separation of N is limited to cell mass and other large molecular forms. Most nitrogen compounds are highly soluble in water.

2 Distillation removal of NH_3 is partially achieved by blow-off of the initial product water vapor or retention of nitrites, nitrates, and ammonia as non-volatile components in the process water. Nitrogen conversion in process makes control difficult.

3 Air stripping of ammonia from a high pH liquid is attractive under certain conditions.

a High concentrations of ammonia strip more readily than low concentrations.

b The air temperature and humidity must be such that stripping is achieved without excessive ice production.

c The high pH requirements favor scale production. Water impurities forming a mud instead of adherent scale are important.

V SUMMARY

A Many methods for removal of N and P are possible. Depending upon the situation, the requirements, concentration, and other factors, one may be favored in one application and not in another.

B Phosphorus removal is inevitably tied to solids removal mechanisms. Good effluent clarification is a necessity to obtain good removal of P. This situation provides a dual function for P removal as many other contaminants are efficiently removed along with phosphorus removal.

C Nitrogen removal is likely to be required in relatively few municipal situations. Nitrogen oxidation to reduce subsequent nitrification demands in surface water is an increasing necessity. A nitrified effluent discharge is likely to result in denitrification by contact with benthic sediments if time is available. The most feasible route for nitrification is a two sludge biological system. If nitrogen removal is necessary, several alternates are possible; all of them require addition of an hydrogen acceptor to speed biological denitrification.

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This outline was prepared by E. F. Barth, Supervisory Research Chemist, Biological Treatment Research Program, Advanced Waste Treatment Research Laboratory, NERC, EPA, Cincinnati, OH 45268.

COMBINED INDUSTRIAL-DOMESTIC WASTE TREATMENT - PAPER MILLS

I INTRODUCTION

A Advantages of combined treatment are varied and numerous, such as: (1)

1 Area planning authorities tend to favor integration into larger units when feasible.

2 The industrial discharge permit system doesn't apply for municipal treatment plants.

3 Construction cost economy approximates the six tenths power rule, for example:

For a single 3 MGD plant in place of 3-1MGD where the 1 MGD plant costs $200,000.

$$\left[\frac{3\text{MGD}}{1\text{MGD}}\right]^{0.6} = \frac{\$x}{\$200,000}$$

$$x = 3^{0.6} \times \$200,000$$

$$= 1.93 \times 200,000 = \$386,000$$

4 It is usually possible to obtain better operational cost/benefit relationships in manpower materials and performance for larger units. Replicate efforts are minimized and continuous control is favored.

5 Separate industrial discharges often have serious nutritional deficiencies that may be eliminated by mixture with other industrial or municipal wastewaters.

6 Monitoring and enforcement commonly are easier with one discharge than with many.

7 Public opposition and land acquisition may be reduced by combined facilities.

8 Combined efforts may be worked out to favor improved financial and personnel adjustments for performance upgrading.

B Certain limitations require careful prearrangement.

1 Close cooperation is required to reduce municipal-industry distrust. Questions related to future commitments or intentions of each body are serious.

2 Financial responsibility of each party for construction, operation and upgrading are complex and difficult to manage.

3 Agreements are necessary regarding what is or is not acceptable for discharge to the treatment facility.

a This usually means an ordinance or agreement regarding the nature of permissible items and limits for flows concentrations or discharge sequence.

b No-no items, such as trash suitable for alternate disposal, explosive or flammable materials, acid or caustic limits, toxic substance limits generally must be specified.

c A mechanism must be provided for handling subsequent changes in discharges, effluent quality control requirements and disagreements. All parties of the agreement and the public need an equitable means of settlement to avoid communication and performance breakdown.

II THE KRAFT PULPING CYCLE (2)

includes the following essentials outlined in Fig 1.

A Raw Materials

1 Wood chips (50% cellulose, 30% lignin, 20% carbohydrates, etc.)

2 Lime ($Ca(OH)_2$)

B Stages

1 Cooking (chips and white liquor) [makes pulp]
2 Chemical recovery (saltcake and black liquor) [makes steam]

3 Liquor making (lime and green liquor) [makes lime sludge]

C Sources of Wastewater

1 Spills (usually black liquor)
2 Washout (of entrained black liquor) from pulp on the paper machine

FIGURE 1.

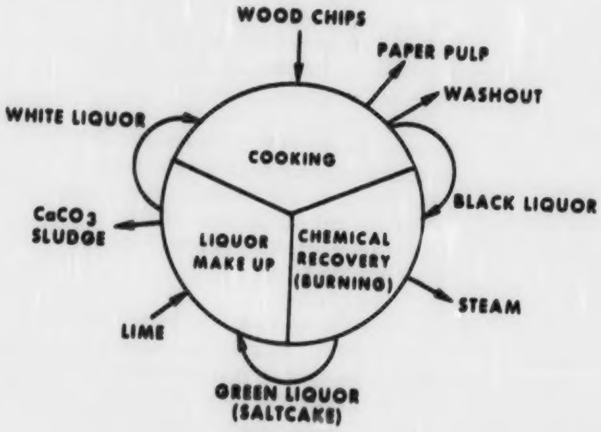

FIGURE 1
KRAFT PULPING CYCLE

III OTHER TYPES OF PULPING AND DIFFERENCES FROM KRAFT PROCESS

A Sulfite

1 Uses sulfuric acid rather than caustic
2 Spent liquor is seldom recovered (Supposed to go to by-products such as road binders for dirt roads).
3 Produces a weaker but more pliable pulp which can make closely-knit, smooth-finished sheets of paper (70% the strength of Kraft pulp).

B Soda

Lye (NaOH) and chips are cooked together. Liquor is often partially recovered. (45% the strength of Kraft pulp).

C Groundwood

1 Wood is ground on a stone with the grain across the width of the stone . . . paralleling its face. (30% of the strength of Kraft pulp).

2 Used for newsprint and other bulk paper requirements where finish strength or tear resistance are not critical.

D Chemi-Groundwood

Four-foot logs dropped in pressure vessels. Soda solution is forced into pores at 150 psi. Soda solution is reused.

IV JOINT TREATMENT AT GREEN BAY, WISCONSIN (3)

A This included municipal wastes and strong wastes from Charmin Paper Products Co. (Ammonia base sulfite pulp process).

1 Trickling filter tests (1958, 1959).

a Single unit and two stage (series) unit. Yeast plant effluent (YPE) spent sulfite liquor (SSL) was used as a substrate for yeast production: BOD reduced by 50%.

b Mixture required more filter capacity per pound of BOD than did municipal wastes alone.

c Idea abandoned as uneconomical.

2 YPE evaporated (10-7% to 55-60%) and burned. (1963).

3 Yeast plant abandoned (1967) in favor of a second boiler for evaporation before burning.

4 Conclusions

a For joint treatment to work best wastes should be compatible to treatment in a conventional plant with conventional loadings and conventional equipment.

b For highly concentrated wastes, various unit operations in chemical engineering usually can be applied more economically than biological treatment.

B Treatment of dilute Charmin Mill wastes with activated sludge. (Evaporator condensate and wash water containing some SSL).

1 No additional nutrients needed.

2 Sludge settling characteristics were poor (Filamentous organisms noted).

C For further study

1 Three more paper mills brought into study for "super plant" design.
2 FWPCA grant obtained.
3 Four activated sludge processes studied.

a Conventional
b Step aeration
c Contact stabilization
d Kraus process

V Brunswick situation: Papermill pulled out of the joint treatment venture.

VI MACON, GA SITUATION (4) INVOLVED THE MUNICIPALITY AND TWO MAJOR INDUSTRIES.

A Pilot plant studies of wastes from

1 City of Macon 24 gpm
2 Armstrong Cork 24gpm
3 Georgia Kraft Co. 72 gpm

The pilot plant loading was proportional to relative volumes from the 3 sources. Solids loading from the Cork installation was highest. The BOD loading from the two industries (4700 and 30,000 lbs/day from 2 and 3 respectively) dominated operation.

B The layout of Pilot Plant is outlined in Fig 2.

C Conclusions

1 Wastes could be treated in combination (90% [BOD] removal).

2 Primary settling needed only for Armstrong Cork wastes.

3 Bulky sludges (1 - 3% solids) were produced, which were difficult to dewater.

4 Supplemental nutrients not needed.

D Subsequent developments

1 Armstrong Cork pulled out (separate plant now built).

2 Thirty-year agreement between Macon and Georgia Kraft for joint treatment. Company has operational control.

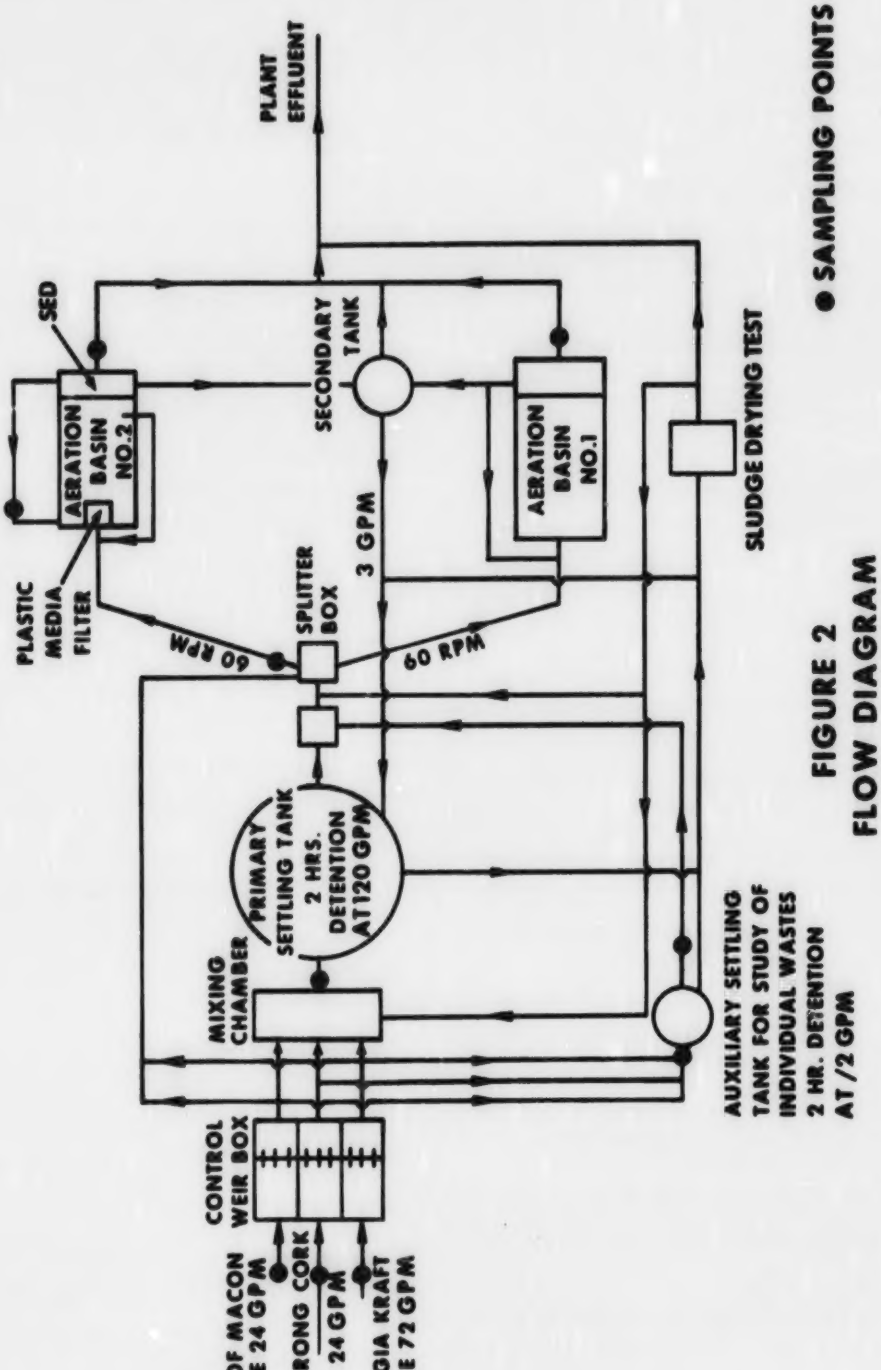
CITY OF MACON WASTE 24 GPM
ARMSTRONG CORK WASTE 24 GPM
GEORGIA KRAFT WASTE 72 GPM
CONTROL WEIR BOX
MIXING CHAMBER
PRIMARY SETTLING TANK 2 HRS. DETENTION AT 120 GPM
SPLITTER BOX
60 RPM
60 RPM
PLASTIC MEDIA FILTER
AERATION BASIN NO.2
SED
AERATION BASIN NO.1
3 GPM
SECONDARY TANK
PLANT EFFLUENT
SLUDGE DRYING TEST
AUXILIARY SETTLING TANK FOR STUDY OF INDIVIDUAL WASTES 2 HR. DETENTION AT /2 GPM
● SAMPLING POINTS

FIGURE 2
FLOW DIAGRAM
PILOT PLANT

ROCKY CREEK WATER POLLUTION CONTROL PLANT

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OXYGEN AERATION FOR SECONDARY TREATMENT

I INTRODUCTION

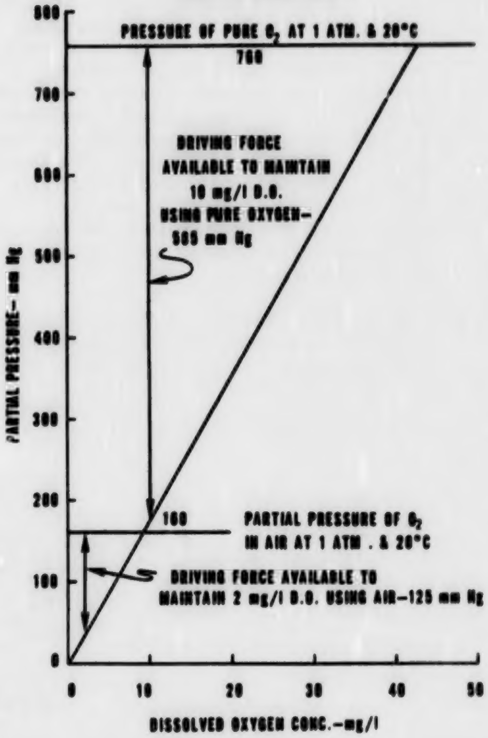
A A new secondary wastewater treatment process has been developed which utilizes oxygen gas, in lieu of atmospheric air, for aeration of activated sludge. The process is called oxygen aeration, or more simply, oxygenation. Development of this process has proceeded rapidly. Today, technology is available for using oxygen aeration in full-scale wastewater treatment plants. Several large oxygenation treatment systems are now under construction and many more are in the design stage.

B Oxygen aeration is an old concept which has received renewed interest. During the past three years, the full transition from drawing board to a marketable process has been made. Completing the cycle from pilot plant evaluation to field application in such a short time is testimony to both the technical and marketing skills of the developers.

C The potential benefits of using oxygen gas in activated sludge treatment have been realized for over twenty years. Oxygen gas, however, is a relatively expensive commodity. Serious consideration of its substitution for air first necessitated the development of a method for efficiently utilizing the input gas. Conventional aeration practices utilize uncovered aeration tanks which result in excessive venting of undissolved oxygen gas, regardless of the particular transfer device selected. The oxygen aeration systems which have been developed recently prevent excessive venting by utilizing a combination of covered or enclosed aerators and gas recirculation. These systems achieve greater than 90% utilization of feed oxygen.

D High partial pressure of oxygen gas has stimulated much interest over the years. As seen in Figure 1, at atmospheric pressure and 20°C, pure oxygen has a partial pressure of 760 mm of mercury compared to 160 mm of mercury for oxygen in air. The available driving force to maintain 10 mg/l of dissolved oxygen (DO) with pure oxygen is 585 mm of mercury. This is approximately 4.7 times larger than the available driving force of 125 mm of mercury to maintain 2 mg/l of DO with air. The system provides a handy reserve of oxygen for peak demands as well as driving force advantages.

FIGURE 1
COMPARISON OF DRIVING FORCES
PURE OXYGEN VS. AIR



II DESCRIPTION OF OXYGENATION SYSTEMS

A Currently, there are three known flow regimes for contacting oxygen with wastewater which are being designed for new treatment systems. They are:

1 Union Carbide System

Covered tank, staged liquid flow, staged gas flow.

2 Cosmodyne System

Covered tank, conventional longitudinal liquid flow, completely mixed gas flow.

3 AIRCO System

Pipeline reactor, ideal liquid and gas plug flow.

These systems are described below:

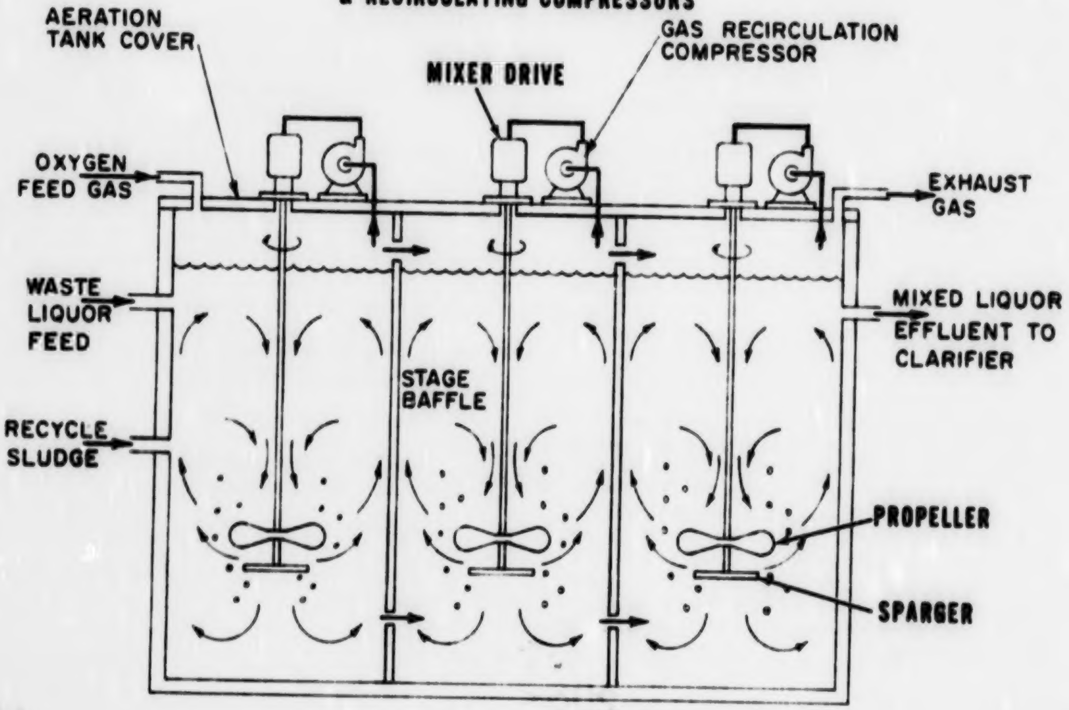
B Union Carbide System (UNOX)

1 Union Carbide rekindled interest in the large-scale application of oxygen in wastewater treatment with the development of their multi-stage system. This system, now marketed under the trade name "UNOX," was initially demonstrated in 1969 under EPA sponsorship at the Batavia, New York Municipal Treatment Plant. The technical success of this demonstration project is largely responsible for generating the necessary stimulus for continued development and commercialization.

2 The "UNOX" system is designed to function in a conventional plug flow configuration. A gas-tight cover is fitted to the top of the aeration tank or poured integrally with the aeration tank to prevent unrestricted loss of oxygen gas to the atmosphere. Vertical baffles are installed at regular intervals, to form a series of compartments or stages. The number of stages for any given system is determined by the length of the aeration tank. To achieve efficient mixing, each stage is sized so that the compartment formed is roughly square.

3 Union Carbide has developed two techniques for dissolving oxygen gas within this multi-stage concept; one utilizes a submerged propeller-sparger combination and a recirculating gas compressor in each stage, the other relies solely on a surface aerator in each stage. A typical three-stage "UNOX" system equipped with propeller-spargers and recirculation compressors and of the type demonstrated at Batavia is shown in Figure 2.

FIGURE 2
SCHEMATIC DIAGRAM OF MULTI-STAGE OXYGENATION SYSTEM UTILIZING TURBINE-SPARGERS & RECIRCULATING COMPRESSORS


AERATION TANK COVER
GAS RECIRCULATION COMPRESSOR
MIXER DRIVE
OXYGEN FEED GAS
EXHAUST GAS
WASTE LIQUOR FEED
MIXED LIQUOR EFFLUENT TO CLARIFIER
STAGE BAFFLE
RECYCLE SLUDGE
PROPELLER
SPARGER

a Gas contained in the vapor space is compressed and forced down the hollow propeller shaft and out through the rotating sparger into the mixed liquor of each stage. The rotating spargers shear the gas bubbles and provide fluid mixing. The recirculating gas compressors can be mounted on top of the tank cover or to the side of the aeration tank, space permitting.

b Oxygen gas is fed into the system at one point only, the lead stage. The inlet gas is 100% oxygen if supplied from liquor storage or by a cryogenic generator. The purity drops to about 90% for a pressure-swing adsorption molecular sieve generator. The rate of oxygen gas feed is controlled by maintaining a positive pressure of 2-3 inches of water column in the vapor space. A control valve automatically permits the oxygen feed rate increase or decrease with demand rate until the desired pressure is re-established.

c Gas and liquid proceed from stage to stage in co-current fashion through appropriately sized openings in the vertical baffles. A portion of the unassimilated oxygen gas that vents back into the vapor space of each stage is continuously recompressed and recirculated through the hollow propeller shaft and out the sparger. The remaining gas is continuously forced through the baffle opening by the 2-3 inch water pressure maintained in the vapor space. In this manner, the vent gas from each stage becomes the feed gas for the succeeding stage.

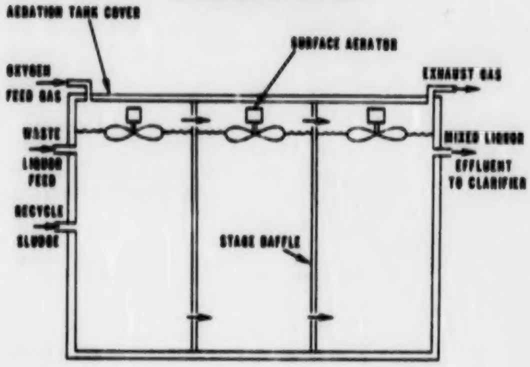
d The vapor space gas becomes increasingly contaminated with gaseous products of decomposition and inert gases in each succeeding stage. By the time the gas flow reaches the last stage, oxygen composition has fallen to approximately 50%. This does not mean that each succeeding compressor must be larger than the preceding one because the net forward flow rate of oxygen gas decreases faster than the drop in oxygen composition. The exhaust gas rate from the last stage to the atmosphere will vary from 10-20% of the incoming oxygen feed rate to the lead stage, which combined with the drop in oxygen composition through the system from 100% to 50% yields an overall oxygen utilization of 90-95%. The positive pressure maintained in the vapor space prevents atmospheric air from backing into the last stage.

e Mixed liquid dissolved oxygen is controlled in each stage by the compressor recirculation rate for that stage. In the lead stage where oxygen demand is greatest and dissolved oxygen levels are more likely to fluctuate, it may be necessary to automatically control compressor recirculation rate with feedback loop tied into a dissolved oxygen sensor. In downstream stages, automatic dissolved oxygen control will normally not be required.

f The multi-stage configuration is also utilized in Union Carbide's surface aerator system (Figure 3). The surface aerators perform the dual function of liquid mixing and oxygen dissolution. Liquid and gas flows proceed co-currently from stage to stage as in the submerged propeller-sparger system. Union Carbide is finding that surface

FIGURE 3

SCHEMATIC DIAGRAM OF MULTI-STAGE OXYGENATION SYSTEM UTILIZING SURFACE AERATORS



aerator systems are favorably cost effective for shallow tank designs and medium strength wastes up to about 50 mgd. The submerged propeller-sparger option is better suited to deep tank geometry. The major design difference in the two alternative systems is the mixed liquor DO operating level. A design DO level of 6 mg/l is usually selected for the surface aerators compared to 8-10 mg/l for the submerged propeller-spargers. This difference is dictated by the greater energy requirement of surface aerators to transfer an equivalent quantity of oxygen gas. For this reason, submerged propeller-sparger designs will generally be preferred for strong wastewaters with high oxygen demands.

C Cosmodyne System (SIMPLOX)

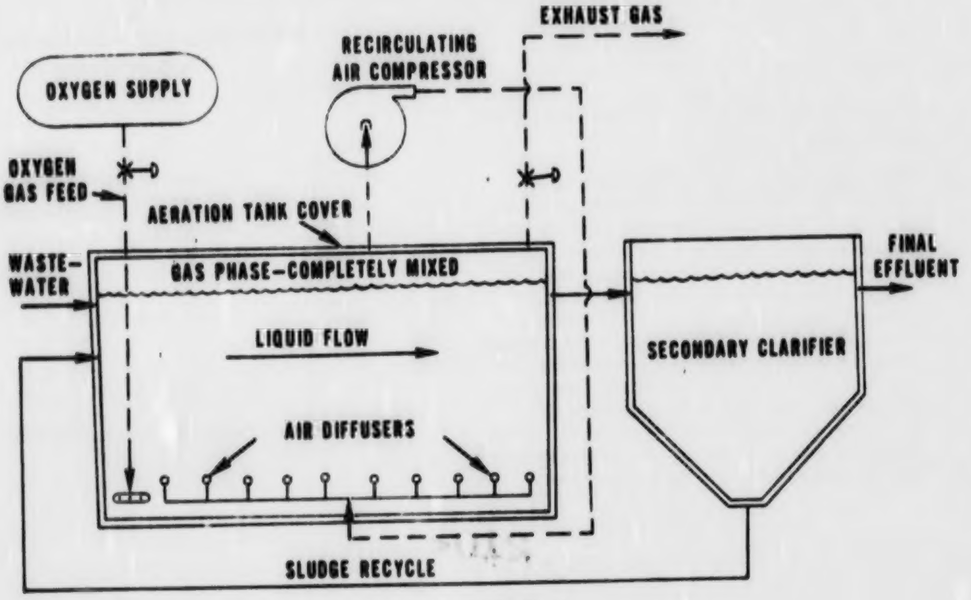
1 An EPA grant has recently been awarded to the Las Virgenes, California Municipal Water District to demonstrate the Cosmodyne Corporation "SIMPLOX" process (Figure 4).

2 The "SIMPLOX" process is intended for application to existing activated sludge plants already equipped with air compressors (of the centrifugal type) and air diffusers. Oxygen gas is introduced through a sparger to the head of the tank. Gas venting into the vapor space is continuously recycled through the air compressors and air diffusers. Exhaust gas is bled from the system at the exit end of the aeration tank.

3 The objective of this project will be to demonstrate that oxygen aeration can be used successfully to increase the organic loading capacity of existing plants with minimum capital expenditure. The additional investment consists of a tank cover, corrosion-proofing of the suction side of the centrifugal air compressors (positive displacement blowers are not suitable) replacing the seals in the air compressors, gas piping, instrumentation and, of course, an oxygen supply source. Staging baffles and multiple stage compressors and propeller-sparger combinations (or surface aerators) are not used in this concept.

FIGURE 4

SCHEMATIC DIAGRAM OF DIFFUSED AIR AERATION SYSTEM MODIFIED TO RECIRCULATE OXYGEN GAS



4 The utilization of existing equipment initially designed for conventional air aeration must be accompanied by a reduction in gas transfer efficiency and an increase in specific power consumption. Gas in the vapor phase will become completely mixed and assume a uniform composition identical to that of the exhaust gas. The available driving force for dissolving oxygen will be smaller than that available in the lead "UNOX" stages where oxygen gas has a lower content of degradation and inert gases.

5 In the final analysis, the trade-offs must be compared: reduced capital costs at the sacrifice of increased operating costs. The developers of the "SIMPLOX" system believe that it will be capable of increasing the organic loading capacity of existing plants by 80% of the incremental increase that is possible with "UNOX."

6 The Las Virgenes project is now in the installation phase. Final project results will be available in about fifteen months. It is likely that this system will eventually be applicable to both conventional plug flow and completely mixed versions of the activated sludge process.

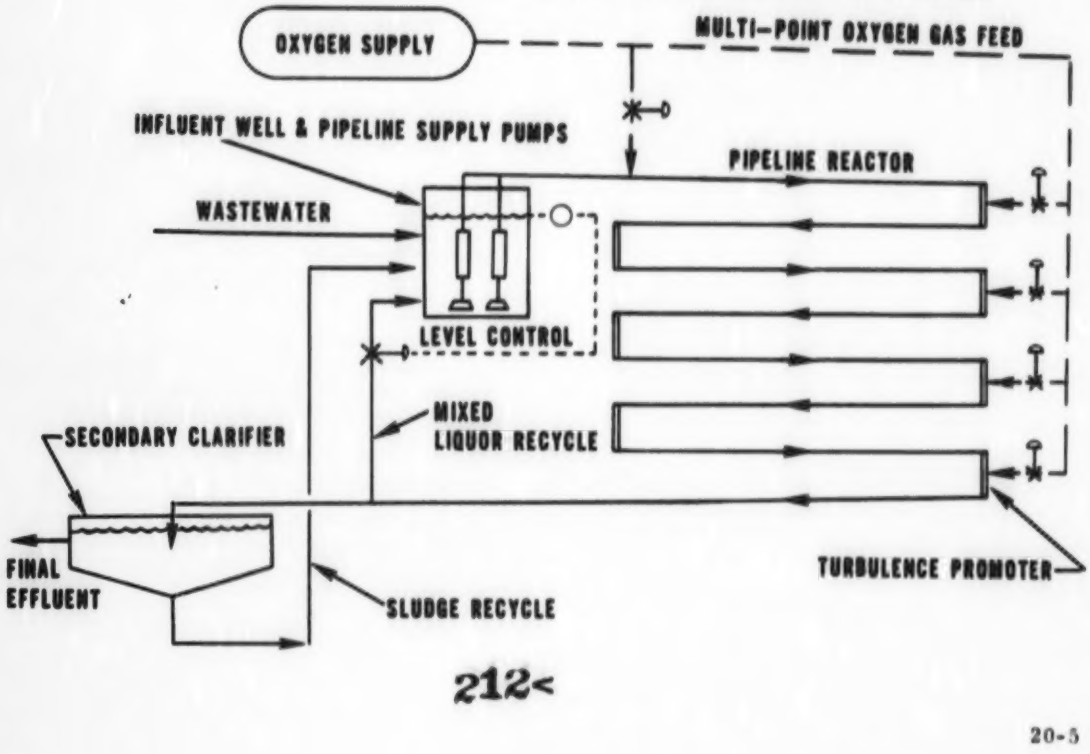
D AIRCO System

1 A completely different type of oxygenation system, the pipeline reactor, has been developed by AIRCO (Figure 5).

2 A large diameter pipeline (30-36") replaces the traditional aeration tank as the biological reactor. Wastewater and oxygen gas move forward in unison in ideal or theoretically perfect plug flow. The only moving parts required are the pumps which supply wastewater and recycle sludge to the pipeline. Direct mixed liquor recycle is also used during low flow periods to maintain

FIGURE 5

SCHEMATIC DIAGRAM OF PIPELINE REACTOR OXYGENATION SYSTEM


OXYGEN SUPPLY
MULTI-POINT OXYGEN GAS FEED
INFLUENT WELL & PIPELINE SUPPLY PUMPS
PIPELINE REACTOR
WASTEWATER
LEVEL CONTROL
MIXED LIQUOR RECYCLE
SECONDARY CLARIFIER
FINAL EFFLUENT
SLUDGE RECYCLE
TURBULENCE PROMOTER

liquid velocity in the pipeline at a minimum of 1 ft/sec.

3 In large diameter pipes, it has been found that velocities of 1-2 ft/sec do not create sufficient turbulence to dissolve any appreciable quantity of oxygen gas. The two materials simply flow down the pipe in two distinct phases, with little interaction. It is, therefore, necessary to create artificial turbulence. AIRCO utilizes vertical U-tubes at points of oxygen injection as turbulence promoters. The U-tubes and multi-gas injection points are staggered along the pipeline at intervals dictated by oxygen demands.

4 A typical AIRCO U-tube oxygen transfer device is illustrated in Figure 6. This particular drawing shows the U-tube extending down into the ground below the mean pipeline elevation. An inverted U-tube which rises above ground can also be used. By injecting oxygen gas into the top of the downward leg a "waterfall effect" is created, thereby setting up violent turbulence and mixing in the "pool" below. A typical 4 ft. head loss will dissolve oxygen to about 50% of saturation. Thus, if pure oxygen is injected, approximately 20 mg/1 of oxygen will be dissolved in each U-tube. The height of the head loss is proportional to the concentration of oxygen dissolved and is controlled by the gas injection rate. Gas injection rate is in turn controlled at each U-tube by utilizing a head loss sensor to maintain a predetermined pressure drop. If more dissolved oxygen is needed to carry to the next U-tube, the head loss set point is increased, and vice versa.

5 The pipeline reactor will find its greatest application at treatment sites where land is not at a premium. It offers several potential construction and operating advantages including:

a A minimum of deep excavation.

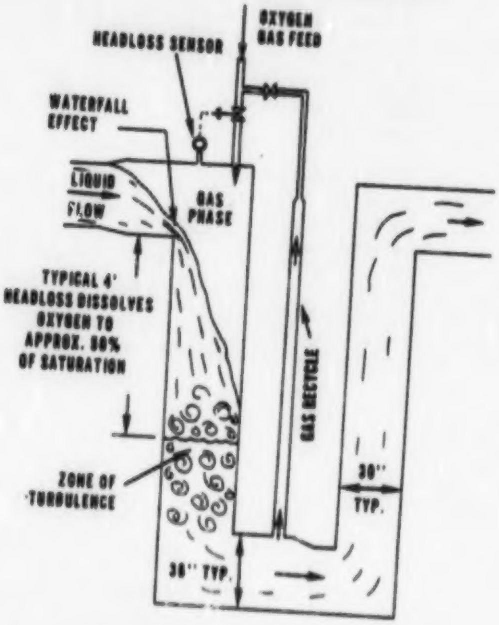
b Rapid installation

c Savings in maintenance costs due to the relative absence of moving parts.

d Easy cleaning (sponge or elastomeric ball).

e Optimum biological kinetics of ideal plug flow.

FIGURE 6
SCHEMATIC DIAGRAM OF U-TUBE OXYGEN TRANSFER DEVICE FOR PIPELINE REACTOR



6 In addition, a covered pipeline has a aesthetic value and it's quite possible that the overlaying land surface may have auxiliary utility. The major disadvantage is anticipated to be locating and repairing leaks in the pipeline if and when they occur. At this time, the capital and operating costs of the AIRCO system are estimated to be roughly competitive with those of "UNOX" where both systems are utilized in new plant construction situations. Cost estimating for upgrading of existing secondary plants must be done on an individual basis.

7 Pipeline reactors should also be attractive alternatives for supplying interim treatment requirements. The first full-scale pipeline reactor system is now being designed by AIRCO for the Passaic County, New Jersey Sewer Authority for installation at

Haskell, New Jersey. A regional plant is expected within 3-4 years but secondary treatment is needed now to prevent raw wastewater discharge into a water supply reservoir. The pipeline will be installed in such a manner that it will become part of a wastewater force main when the regional plant is built.

III THE BATAVIA PROJECT

A The majority of the full-scale performance data subsequently presented were generated during the 1969 Batavia, New York project. A brief description of this project is given below.

1 A Research and Demonstration Contract was awarded by EPA (then FWPCA) to Union Carbide in the fall of 1968 to compare the performance and economics of their newly developed multi-stage oxygenation system with those of an existing conventional plug flow diffused air aeration system. The Batavia treatment plant was selected because it was suited to conducting a parallel process study. The plant is equipped with two parallel mirror-image treatment trains consisting of dual two-pass aeration tanks, dual secondary clarifiers, and separate sludge recycle facilities for each train. Other plant units include screening, grit removal, and comminution facilities, a secondary sludge thickener, a two-stage high-rate anaerobic digestion system, a vacuum filter, and chlorination facilities. Primary clarification studies were not possible at this plant.

2 One of the dual two-pass air aerators was converted to oxygen aeration. Liquid oxygen, stored and vaporized on site, provided the oxygen supply to the oxygen aeration train. The existing conventional plug flow diffused air aeration system was left intact in the second train and was "beefed" up with additional diffusers to permit higher than conventional organic loadings to that train. The nominal design flow of the Batavia plant is 2.5 mgd, 1.25 mgd to each train. The flow pattern utilized during the experimental program is given in Figure 7. It was not possible to separately thicken and digest the waste activated sludge from the two systems. Consequently, the thickener and digester supernatants recycled to the aeration tanks were a mixture from both systems.

B Summary of Performance Data

1 The Batavia project was conducted in three phases. In Phase 1, the raw wastewater flow was split equally between the two trains utilizing the full available aerator capacity of each system. All of the plant wastewater flow was diverted through one half of the oxygenation system in Phase 2. In Phase 3, the wastewater flow was again split between the two trains, but only half of the available aerator volume of each system was used. The data presented in the following tables represent average values for each phase.

2 Aeration detention times and mixed liquor characteristics of the two systems are summarized below in Table 1.

3 The difference in MLSS concentrations held in the two systems was due to oxygen transfer limitations in the spiral flow diffused air system. Approximately 3600 mg/l was the maximum MLSS level possible in the air system at maximum output without sepsis in the biomass.

4 The mixed liquor in both systems exhibited excellent settling characteristics. The very low SVI's for both systems were due, in part, to the absence of primary clarification and the resultant influx of more dense solids into the mixed liquor.

5 Predictable effect of a closed system on pH is readily apparent in Table 1. The cover on the Union Carbide oxygenation system prevents the normal stripping of carbon dioxide to the atmosphere associated with all open tank air aeration systems. Consequently, a much larger fraction of the carbon dioxide produced in substrate oxidation remains dissolved in the mixed liquor, thereby, lowering pH. The drop in pH in the oxygenation train was not drastic, ranging from 0.5-0.7 pH units lower than the air aeration train.

WATER POLLUTION CONTROL PLANT, CITY OF BATAVIA, N.Y.
SCHEMATIC FLOW DIAGRAM

FIGURE 7

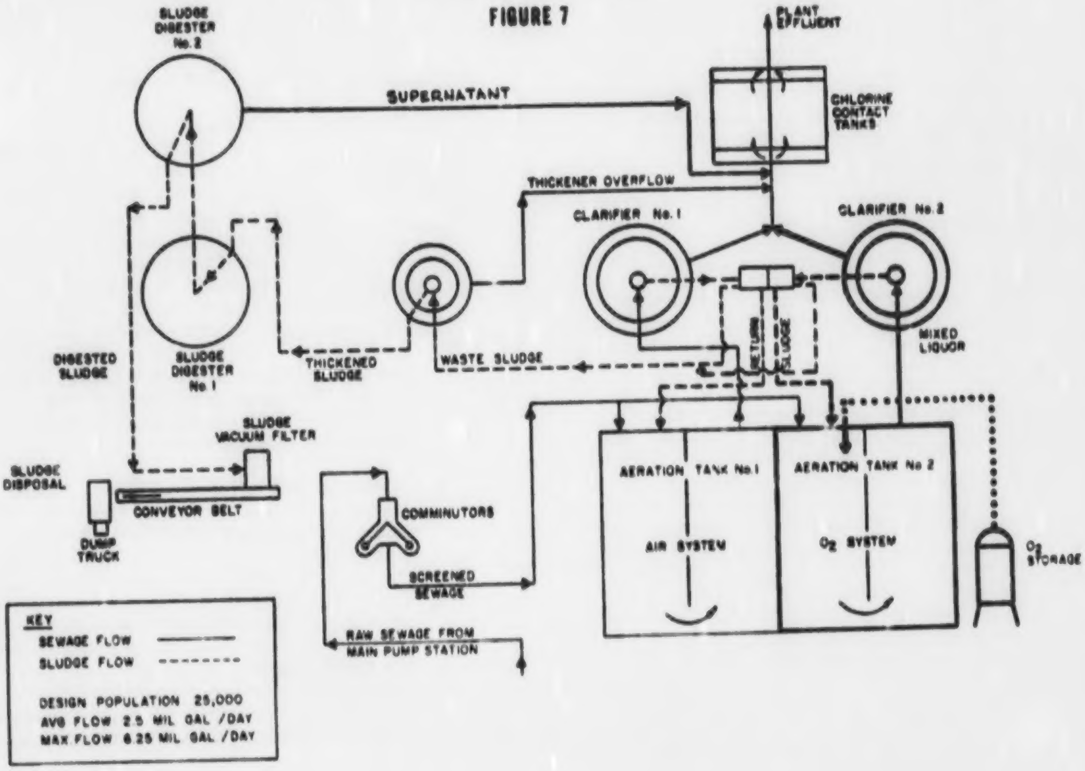

SLUDGE DIGESTER No. 2
PLANT EFFLUENT
SUPERNATANT
CHLORINE CONTACT TANKS
THICKENER OVERFLOW
CLARIFIER No. 1
CLARIFIER No. 2
DIGESTED SLUDGE
SLUDGE DIGESTER No. 1
THICKENED SLUDGE
WASTE SLUDGE
RETURN SLUDGE
MIXED LIQUOR
SLUDGE VACUUM FILTER
SLUDGE DISPOSAL
CONVEYOR BELT
DUMP TRUCK
COMMINUTORS
SCREENED SEWAGE
AERATION TANK No. 1
AIR SYSTEM
AERATION TANK No. 2
O2 SYSTEM
O2 STORAGE
RAW SEWAGE FROM MAIN PUMP STATION
KEY
SEWAGE FLOW
SLUDGE FLOW
DESIGN POPULATION 25,000
AVG FLOW 2.5 MIL GAL /DAY
MAX FLOW 6.25 MIL GAL /DAY

Table 1

	Phase 1		Phase 2	Phase 3	
	Air	Oxygen	Oxygen	Air	Oxygen
Nominal Aeration Time (hr) Raw Flow Only	4.0	4.1	1.5	3.0	2.8
Actual Aeration Time (hr) Raw + Recycle	3.5	3.4	1.2	2.6	2.0
MLSS (mg/l)	2440	3060	6980	3640	6190
% Volatile	71	72	64	71	70
SVI	76	64	36	63	49
pH	7.1	6.4	7.0	7.6	7.1
Mixed Liquor DO (mg/l)	1.5	8.7	9.0	0.8	8.0
Initial Settling Velocity (ft/hr)	7.7	7.2	6.5	7.7	5.1
Wastewater Temperature Range (°F)	53 - 67		60 - 72	66 - 73	

Combining the aeration detention time and MLVSS concentrations indicated in Table 1 yielded the system loading rates of Table 2.

Table 2

	Phase 1		Phase 2	Phase 3	
	Air	Oxygen	Oxygen	Air	Oxygen
Volumetric Organic Loading (lb. BOD/day/1,000 ft^3)	60	58	212	129	145
Food-to-Biomass Loading (lb. BOD/day/lb. MLVSS)	0.57	0.41	0.79	0.84	0.55
Clarifier Surface Overflow Rate (gpd/ft^2)	1570	1420	1010	1030	1140

6 As seen in Table 2, the oxygenation system was able to absorb increased volumetric organic loadings while still retaining reasonable food-to-biomass loadings much better than the diffused air aeration system. The volumetric organic loading imposed on the oxygenation system in Phase 2 (212 lb. BOD/lb/1,000 ft^3) was more than four times the recommended design level for conventional systems. The food-to-biomass loading of 0.79 lb. BOD/day/lb. MLVSS was only slightly higher than normally recommended values. The ability of the oxygenation system to sustain high MLVSS levels under high oxygen demand conditions is responsible for this system characteristic.

7 Only minor problems in liquid-solids separation were encountered in either system despite the high clarifier surface overflow rates experienced throughout the project.

8 Treatment performance for the two systems, as measured by BOD, COD, and suspended solids removals across the aerators and secondary clarifiers, is compared in Table 3.

Table 3

	Phase 1		Phase 2	Phase 3	
	Air	Oxygen	Oxygen	Air	Oxygen
BOD In (mg/l)	159	159	220	262	262
BOD Out (mg/l)	16	11	23	30	14
% BOD Removed	90	92	90	88	94
COD In (mg/l)	352	352	325	578	578
COD Out (mg/l)	84	73	97	116	89
% COD Removed	76	80	71	79	84
SS In (mg/l)	221	221	174	430	430
SS Out (mg/l)	16	9	19	23	12
% SS Removed	93	96	89	94	97

Considering the range of volumetric loadings, both systems performed very well. Oxygenation system performance was, however, superior to that of the air aeration system in both comparative phases. The data also indicate a slightly better performance by the oxygenation system under the highly loaded conditions of Phase 2 than by the air system in Phase 3. Using the above data, a correlation was derived between aerator detention time and BOD removal which indicates that equivalent treatment is provided by an oxygen aerator having one-third the volume of a conventional diffused air aerator. This correlation is seen in Table 4.

Table 4

Nominal Aeration Time (hr-based on raw flow) Oxygen	% BOD Removal Across Secondary System	Nominal Aeration Time (hr-based on raw flow) Air
1.00	84	3.0
1.33	87	4.0
1.67	90	5.0
2.00	92	6.0

9 The Batavia sludge wasting data summarized in Table 5 indicate a substantial reduction in excess biological sludge formed by an oxygenation system.

Table 5

	Phase 1		Phase 2	Phase 3	
	Air	Oxygen	Oxygen	Air	Oxygen
Solids Wastes (lb/day)	3148	1804	2448	3494	592
Lb VSS Wastes/lb BOD Removal	0.87	0.48	0.41	0.99	0.13

a These data and similar data from a follow-on 1970 Batavia project were plotted against BOD removed using MLVSS as a common denominator in Figure 8. The oxygen data from Phase 3 in Table 5 were not used in deriving Figure 8 because they did not appear to be realistic. A 35-45% reduction in sludge production is projected for oxygenation by Figure 8 over the range shown.

b Sludge production data have also been collected on pilot oxygenation and air aeration units located at the EPA Washington, DC Blue Plains Pilot Plant. These data are shown in Figure 9 as a function of Sludge Retention Time (SRT). For SRT's greater than six days, similar reductions in the quantity of excess sludge generated by oxygenation are indicated.

c The sludge wasting data graphically depicted in Figures 8 and 9 have significant potential economic implications, particularly for plants that forego primary clarification and do not have to contend with primary sludge disposal.

d Oxygen supplied and percent oxygen utilized in the three phases are shown in Table 6.

10 Verification of precontract assumptions that an oxygen utilization of at least 90% could be routinely achieved was a vital element in demonstrating a cost effective system. Oxygen supply

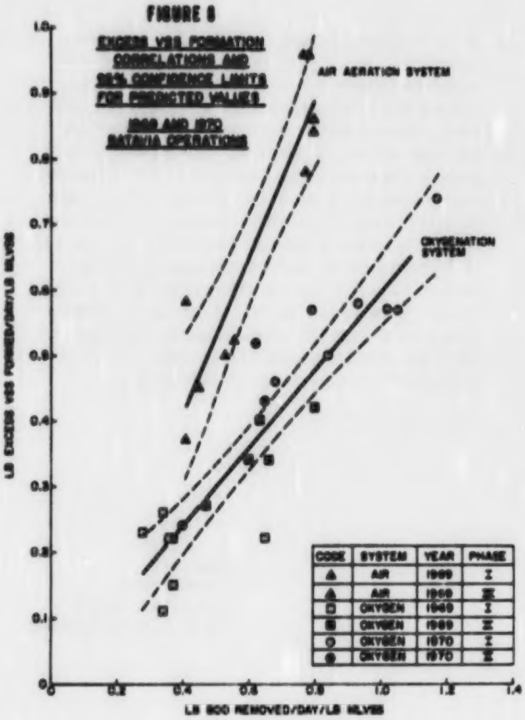


FIGURE 8

EXCESS VSS FORMATION CORRELATIONS AND 95% CONFIDENCE LIMITS FOR PREDICTED VALUES

1969 AND 1970 BATAVIA OPERATIONS

requirements were less than 1.0 lb/lb BOD removed for all phases. Data from other pilot operations indicate that this value may sometimes be greater than unity, particularly at high SRT's.

a Power requirements from the Batavia project are summarized in Table 7.

b As shown, 5-6 times more power was needed for air compression.

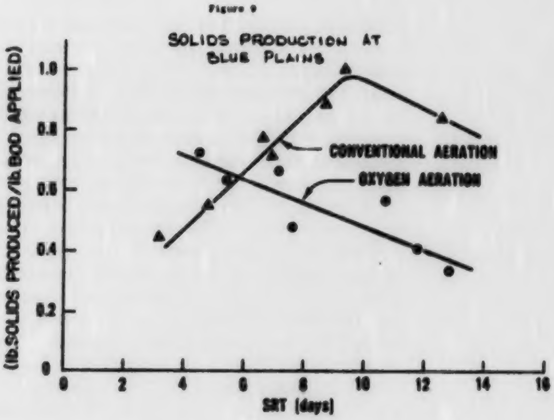


Figure 9

SOLIDS PRODUCTION AT BLUE PLAINS

Table 6

	Phase 1 Oxygen	Phase 2 Oxygen	Phase 3 Oxygen
O_2 Supplied (lb. O_2/ lb. BOD Removed	0.94	0.96	0.74
% O_2 Utilized	95.5	92.7	91.4

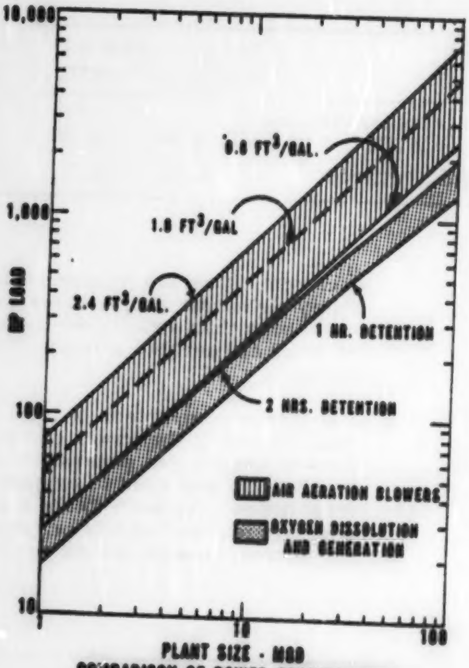
However, it should be noted that much higher air supply rates were necessary than usually encountered at other air aerated activated sludge plants.

c For a true comparison of total power requirements, the power consumed in on-site oxygen gas generation should be added to that used in oxygen dissolution. No such totals were possible for the Batavia project. In general, though, based on Union

Table 7

	Phase 1		Phase 2	Phase 3	
	Air	Oxygen	Oxygen	Air	Oxygen
O_2 Dissolution Power (Total HP utilized)	-	28.6	20.3	-	22.3
O_2 Dissolution Power (HP/1000 gal. mixed liquor)	-	0.088	0.125	-	0.137
Air Blower Power (Total HP utilized)	141	-	-	138.5	

Carbide estimates, power requirements for oxygen generation will range from 1.3-1.7 times greater than for oxygen dissolution, depending on plant size. A band of estimated total power requirements for air compression versus oxygen generation plus oxygen dissolution are shown in Figure 10 for plants from 1 to 100 mgd capacity. The band for air compression corresponds to 0.8-2.4 ft^3 of air/gallon of wastewater treated. The band for the oxygenation system represents a range of 1-2 hours of aeration detention time, based on raw wastewater flow. Using the median level in each band, a total power savings of approximately 50-67% can be projected over this range of plant sizes with oxygenation.



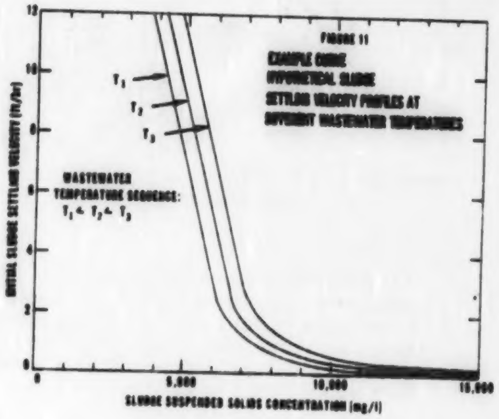
COMPARISON OF POWER REQUIREMENTS

FIGURE 10

IV SECONDARY CLARIFIER DESIGN

A The design criteria which should be adhered to in sizing oxygenation system secondary clarifiers are not yet precisely known. This matter is under study and hopefully definite guidelines can be established in the near future.

B It is believed that eventually oxygen clarifier sizing will be controlled by cold weather conditions. First it is necessary to establish what effect, if any, wastewater temperature has on sludge settling velocity profiles from batch flux settling tests run at different temperatures can aid designers in the proper selection of allowable clarifier mass loadings and surface overflow rates. A hypothetical set of sludge settling velocity profiles is illustrated in Figure 11. Until data from several studies now in progress can be correlated, designers will probably continue to use average overflow rates of 700-800 gpd/ft^2.



V SLUDGE HANDLING

A In addition to generating a smaller mass of excess biological sludge, oxygenation systems appear to produce a more concentrated waste sludge than conventional systems. Waste solids concentrations of 2-3% have been routinely produced at Batavia and other pilot studies around the country, compared to the 1.0-1.5% more typical of air aeration plants.

1 The excellent thickening accomplished in oxygen clarifiers suggests the feasibility of direct vacuum filtration of waste sludge without an interim separate thickening step. To test this hypothesis, pilot-scale vacuum filtration

runs were conducted directly on clarifier underflow sludge as part of the 1970 follow-on Batavia project. Filter cake yields in excess of 4 lb/hr/ft^2 were achieved with 2.0-2.4% feed solids concentrations and cycle times of about 2-1/2 minutes. The solids content of the cake at these high yields averaged 17-18%. A correlation between cake yield, cycle time, and percent feed solids developed by Union Carbide is shown graphically in Figure 12. This correlation equation was applied to City of Milwaukee vacuum filtration data generated on thickened air aerated waste activated sludge. From the slope of the plot in Figure 13, it appears that if unthickened waste oxygenated sludge were filtered under the same conditions as practiced at Milwaukee, an average 43% higher yield would be possible.

2 Future economic benefits are definitely indicated in the handling and dewatering of oxygenated sludges.

FIGURE 12

OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE

OBSERVED CAKE YIELD VS. $\left(\frac{C_f^{0.787}}{t^{0.586}}\right)$

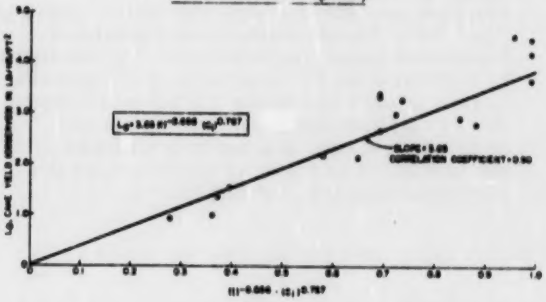
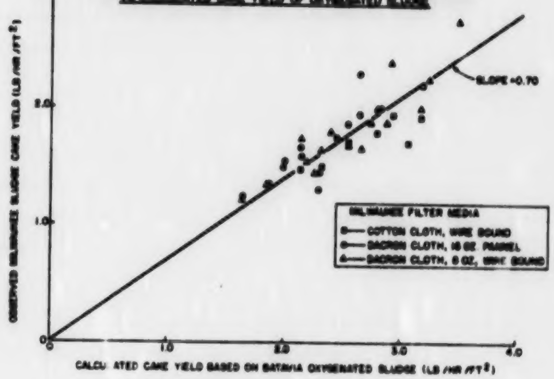


FIGURE 13

OBSERVED CAKE YIELD FROM VACUUM FILTRATION OF MILWAUKEE'S CONVENTIONAL THICKENED WASTE ACTIVATED SLUDGE VS CALCULATED CAKE YIELD OF OXYGENATED SLUDGE



VI SUPPORTIVE PILOT PLANT DATA

Performance data from three Union Carbide mobile pilot plant programs plus data from the Blue Plains operation and the Batavia 1970 follow-on project are summarized in Table 8. These data provide further evidence

Table 8

PILOT PLANT OXYGEN AERATION DATA

Pilot Plant	Aeration** Detention Time (hr)	Volumetric Organic Loading $\left(\frac{\text{lb. BOD/day}}{1000 \text{ ft}^3}\right)$	F/M Ratio $\left(\frac{\text{lb. BOD/day}}{\text{lb. MLVSS}}\right)$	BOD In (mg/l)	BOD Out (mg/l)	% BOD Removal Across O_2 System
New Orleans	1.3-2.1	145-256	0.61-1.00	168-254	9-15	92-96
Cedar Rapids	1.1-4.4	180-500	0.55-1.85	154-425	8-55	87-98
Louisville	1.1-1.8	173-186	0.75-1.00	126-193	19-23	83-89
Blue Plains	1.6-2.5	57-92	0.28-0.39	87-134	11-32* (2-3 Sol.)	76-91
Batavia (2nd year)	1.2-1.6	73-272	0.45-1.23	97-277	27-51* (7-16 Sol.)	60-88

**Based on Raw Flow Only.

*Clarifier Hydraulically Overloaded.

that the oxygen aeration process produces a high quality secondary effluent with short aeration periods and high volumetric loadings. The Cedar Rapids wastewater contains a significant industrial fraction. The relatively high effluent BOD's experienced on occasion at Blue Plains and Batavia (2nd year) were due to clarifier hydraulic overloads and solids carryover, and not to breakdown of the biological process as evidenced by the corresponding low soluble BOD's.

VII LARGE-SCALE PLANT DESIGNS

A Oxygen aeration technology is now being utilized in the design of large-scale wastewater treatment plants. Design criteria have been formulated from the contract and pilot plant programs previously described. Design parameters for two systems presently under construction, Detroit and the Newtown Creek Plant in Brooklyn, are compared in Table 9. A new 300-mgd oxygen aeration module is being installed in Detroit. At the Newtown Creek Plant, an EPA Research and Demonstration Grant is partially supporting the conversion of one existing 20-mgd high-rate activated sludge bay to oxygen aeration.

Table 9

Detroit	Parameter	Newtown Creek
Primary Effluent	Type of Feed	Raw Wastewater
300 mgd	Design Flow	20 mgd
25-33%	Return Sludge Flow	21-30%
140 mg/1	Oxygenator Influent BOD	200 mg/1
25 mg/1	Final Effluent BOD	20 mg/1
5000 mg/1	MLVSS	4500 mg/1
1.42 hr	Aeration Time (Raw Flow)	1.5 hr
1600 gpd/ft^2	Clarifier Overflow Rate	900 gpd/ft$_2$

B The overflow rate in the Brooklyn plant is slightly higher than desired but is set by existing available clarifier surface area. At Detroit, a relatively new type of rim-fed, rim-takeoff clarifier with a very high design overflow rate was selected. This overflow rate is approximately twice that normally used and definitely would not be recommended for center-feed clarifiers.

VIII ECONOMIC PROJECTIONS

A Data from the two Batavia projects were drawn upon to develop comparative total treatment cost estimates for new plants with a multi-stage oxygenation system of Union Carbide design (with submerged propeller-spargers and recirculating compressors) and a conventional diffused air aeration system. Since the majority of the plants being built in this country incorporate primary sedimentation, the cost of this unit process is included in the estimates.

1 Comparative cost estimates were prepared for four sizes of plants, 1, 6, 30, and 100 mgd. These estimates were reduced to total treatment costs in cents per thousand gallons treated, as shown in the curves on Figures 14 through 17.

2 Due to the variance in oxygen transfer capability of different types of air diffusers, curves for three air aeration rates, 0.8, 1.6, and 2.4 ft^3/gallon, were included. The curves reflect both capital debt retirement and operating and maintenance costs for all plant facilities and sludge disposal equipment, except effluent chlorination. Costs were determined using January, 1970 indices.

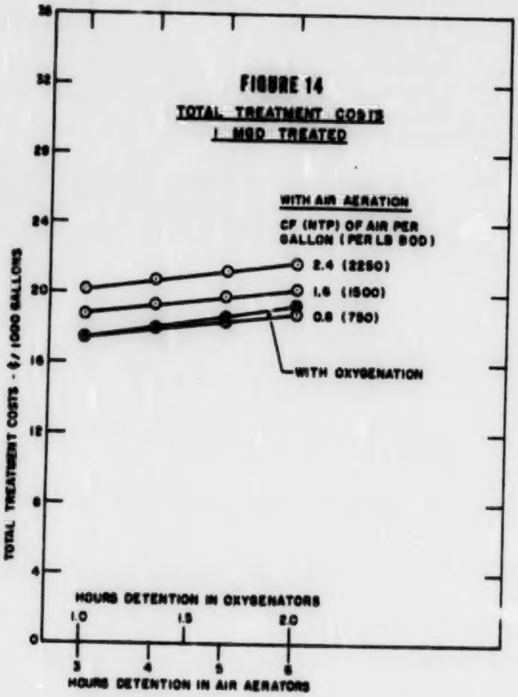


FIGURE 14
TOTAL TREATMENT COSTS
1 MGD TREATED

3 From these four figures, it can be seen that there is a significant cost advantage with oxygenation for all plant sizes estimates except 1 mgd. At 1 mgd, there is no apparent cost advantage for either system. Using the lower air curves of 0.8 ft^3/gallon, at the 6, 30, and 100 mgd plant scales, estimated cost savings of 8-12%, 15-18%, and 12-17%, respectively, are projected with oxygenation. For the middle air curves of 1.6 ft^3/gallon, the projected savings increase to 14-17%, 20-22%, respectively.

Figures 15, 16, and 17 are expanded into one curve in Figure 18 to show anticipated ranges in total treatment costs from 6-100 mgd.

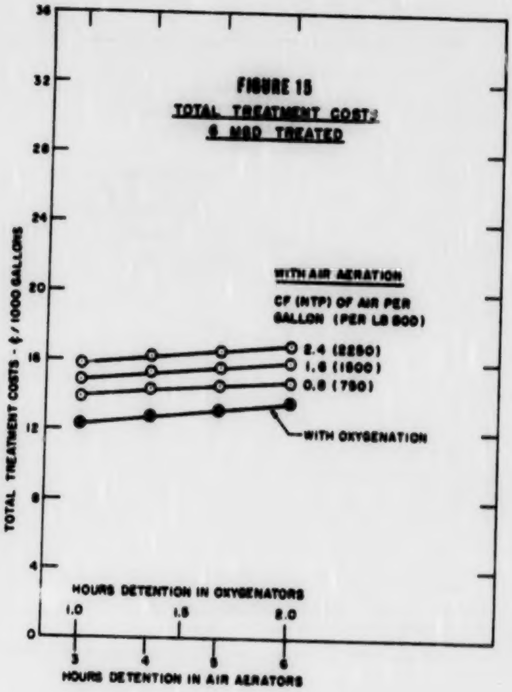


FIGURE 15
TOTAL TREATMENT COSTS
6 MGD TREATED

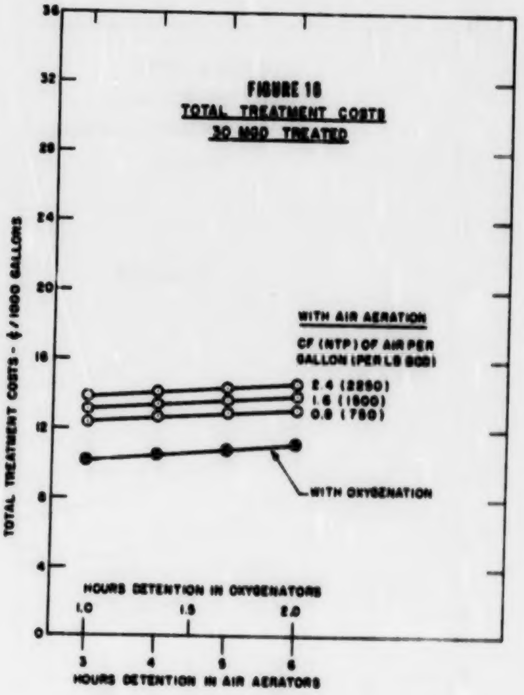


FIGURE 16
TOTAL TREATMENT COSTS
30 MGD TREATED

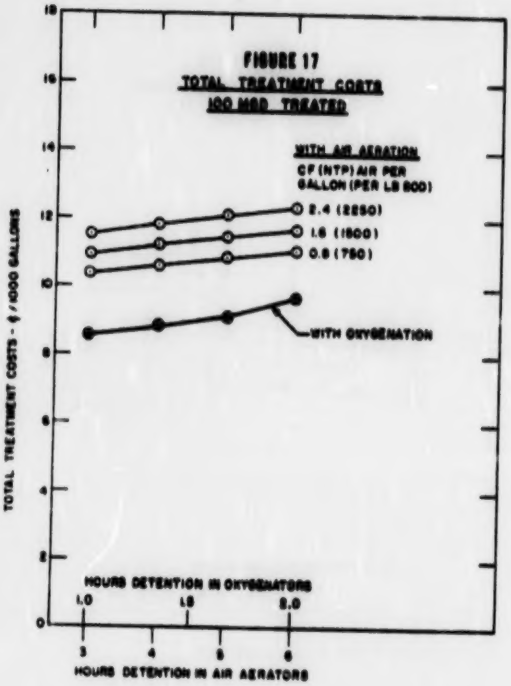
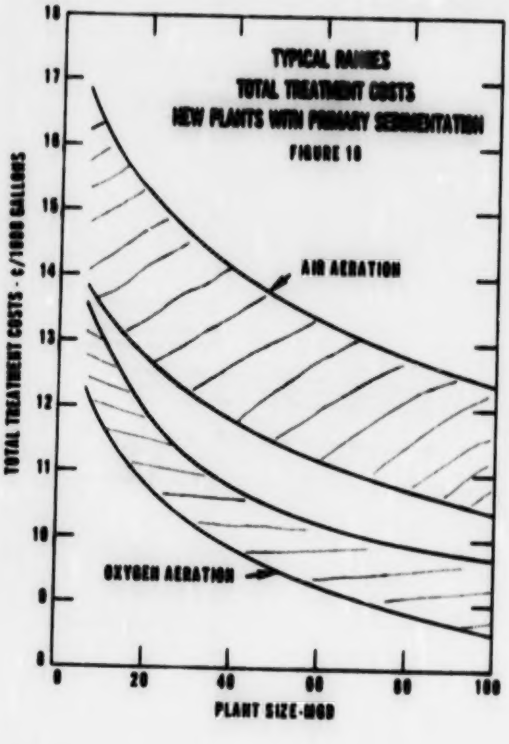


FIGURE 17
TOTAL TREATMENT COSTS
100 MGD TREATED


TYPICAL RANGES
TOTAL TREATMENT COSTS
NEW PLANTS WITH PRIMARY SEDIMENTATION
FIGURE 18

It is emphasized that the cost data shown in the preceding figures are best estimates only and are based on assumptions that may not apply to every situation. Individual plant costs will vary widely with locale and construction conditions. Further, cost estimates may vary significantly for air aeration systems which utilize mechanical aerators or modifications of the activated sludge process other than conventional plug flow. The assumptions and methods used in conjunction with the Batavia data in preparing the comparative cost curves can be found in Water Pollution Control Research Series Report 17050 DNW 05/70.

4 The major factors responsible for the estimated reduced treatment costs with oxygenation are presented in Table 10.

Table 10

MAJOR FACTORS RESPONSIBLE FOR PROJECTED REDUCED TREATMENT COSTS WITH OXYGENATION

1 2-3 fold decrease in required aerator volume

2 Approximate 50% reduction in aeration power requirements

3 Approximate 1/3 reduction in mass of excess biological sludge produced

4 Improved thickening and handling characteristics of excess biological sludge

Major investment items are listed in Table 11 indicating which system, oxygen or air, will incur the greater investment cost for each item.

5 A major operating cost for an oxygenation system is the purchase cost of the oxygen or the cost of the power required to generate it on-site. The various options open to a municipality in securing an oxygen supply are indicated in Figure 19. As shown, the unit cost is much cheaper if an on-site generation plant is purchased directly from a manufacturer and operated by municipal personnel. In making such a decision, though,

Table 11

TOTAL PLANT INVESTMENT LIST COMPARISON
OXYGENATION VS. AIR AERATION

Item	Investment Cost Greater for	
	O_2	Air
Aeration Tank		X
Tank Cover and Stage Baffles	X	
O_2 Dissolution + Generation Equipment vs. Air Blowers	X	
Sludge Return Equipment		X
Sludge Digesters		X
Sludge Thickening and Digestion Facilities		X
Land		X
Secondary Clarifiers	Same	
Preliminary and Primary Treatment Facilities	Same	
TOTAL		X*

*Applies to plants 5 mgd or larger.

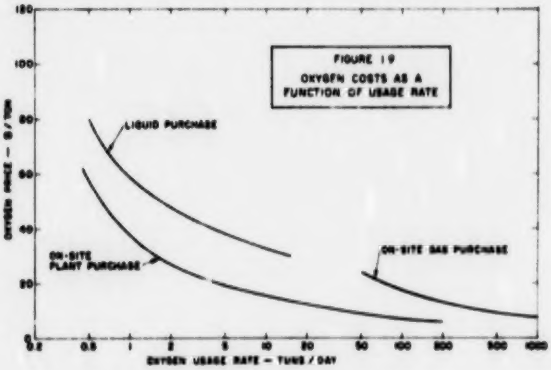


FIGURE 19
OXYGEN COSTS AS A FUNCTION OF USAGE RATE

the municipality must also consider the initial investment involved.

IX OXYGEN AERATION - THE PRESENT

A Several factors have been fairly well established over the past three years. They include among others that:

1 Oxygenation systems posses excellent oxygen transfer and utilization characteristics which result in significant reductions in dissolution power requirements.

2 Oxygenation is a highly oxidative process which converts a greater fraction of organic substrates to carbon dioxide and water than a conventional air aerated activated sludge system, thereby resulting in decreased biological sludge production.

3 Oxygenation systems can simultaneously support high MLSS concentrations (6,000-7,000 mg/l) and maintain a mixed liquor DO level of 8-10 mg/l, and are, therefore, capable of operating effectively under very high volumetric organic loadings and at short aeration detention times.

B Sufficient confidence has been generated in the process that today ten municipal plants totaling more than 600 mgd of flow are in various stages of design and construction. These plants are listed in Table 12. In addition, another fourteen municipal plants totaling another 600 mgd will soon go to the drawing boards.

Table 12

LARGE SCALE MUNICIPAL OXYGEN AERATION PLANTS BEING DESIGNED OR UNDER CONSTRUCTION

Plant	Flow (mgd)
Detroit, Michigan	300
Brooklyn, New York	20
Speedway, Indiana	8
Winnepeg, Canada	12
New Rochelle, New York	14
Middlesex, New Jersey	120
New Orleans, Louisiana	122
Fairfax County, Virginia	14
Passaic County, New Jersey (Pipeline Reactor)	4
Las Virgenes, California	2

C Oxygenation has also made in-roads into the industrial waste treatment field. The three industrial applications shown in Table 13 will utilize oxygen for treatment of high-strength wastes.

Table 13

OXYGEN AERATION PLANTS BEING DESIGNED OR UNDER CONSTRUCTION FOR INDUSTRIAL APPLICATIONS

Plant	Flow (mgd)
Pearl River, New York (Pharmaceutical BOD-1600)	1.5
Peekskill, New York (Fermentation BOD-960)	1.0
Taft, Louisiana (Petrochemical BOD-2600)	3.6

X OXYGEN AERATION - THE FUTURE

A It is believed that oxygen aeration will find extensive application in the future in three major areas:

1 Upgrading of existing overloaded activated sludge plants by conversion from air aeration to oxygen aeration.

2 Upgrading of existing trickling filter plants by adding oxygen aeration as a second stage biological step.

3 New plant construction both with and without primary sedimentation.

B Potential "spin-off" applications that are quite probable, particularly where oxygen gas is generated on-site, include:

1 Post-oxygenation

2 Aerobic sludge digestion with oxygen

3 Sludge incineration with oxygen

This outline was prepared by Richard C. Brenner, Research Sanitary Engineer, Biological Treatment Research Program, AWTRL, RATWRC, NERC, Environmental Protection Agency, Cincinnati, OH 45268.

CHAPTER 3

FLOW EQUALIZATION

3.1 General

The cyclic nature of wastewater flows, in terms of volume and strength, is well established. While the concept of flow equalization has been employed in the field of water supply and in the treatment of some industrial wastes, it has not been widely accepted in the municipal pollution control field. Anticipated problems with solids settling, odor, and septicity can be cited as the major factors limiting its use.

Recently, interest in flow equalization for municipal treatment has increased due to the advent of stricter water quality standards, the elimination of plant bypassing, and the increased removal efficiencies that are possible when biological or chemical treatment processes are operated at or near steady-state conditions.

There are two major objectives in the design of flow equalization basins. The first of these is simply to dampen the diurnal flow variations that normally exist in typical municipal wastewater collection systems, and thus achieve a constant or nearly constant flow rate through the downstream treatment processes. In this type of system, little consideration is given to controlling the concentration changes that take place during storage. The major design factors are supplying enough air to keep the basin aerobic and providing adequate mixing to prevent solids deposition. Consideration should be given to locating equalization basins both at the treatment plant site and at strategic upstream locations in the tributary collection system. The upstream locations may offer the added advantage of relieving trunk sewer overload during peak flow periods.

The second objective of flow equalization is to provide the capacity to distribute shock loads of toxic or treatment-inhibiting substances over a reasonable period of time to prevent system failure and to minimize the periodic discharge of harmful contaminants to the receiving stream or surface impoundment. The measurement or estimation of time-dependent concentration profiles and flow-through curves is normally used to analyze the flow characteristics of these systems for determining the effects of tank geometry, effluent weir placement, and mixing regime on changes in contaminant concentrations through the basin.

In all cases, the added costs of flow equalization must be measured against the reduction in downstream process costs and the increased efficiencies that can be achieved by operating these processes under relatively constant loading conditions.

3.2 Determination of Equalization Requirements

To determine the appropriate equalization basin volume, it is necessary to plot an inflow mass diagram of the hourly fluctuations for a typical daily wet-weather wastewater flow.

This copy taken from the "Process Design Manual for Upgrading Existing Wastewater Treatment Plants" prepared by Roy F. Weston, Inc. under Contract 14-12-933 with the EPA Technology Transfer Program.

Figure 3-1 shows the hourly fluctuations for a typical plant. Superimposed on Figure 3-1 is the inflow mass diagram for the hourly flows, the ordinate of which is obtained by accumulating the hourly flows and converting them into equivalent volumes of wastewater.

In Figure 3-1, the slope of line A represents the average rate at which the wastewater is pumped from the equalization basin to the downstream treatment units, which for the particular wet-weather flow in Figure 3-1 is 10,000 gallons/hour. This slope is determined by drawing a straight line through the origin and point C, which is the end of the inflow mass diagram. The maximum required capacity of the equalization basin is determined by drawing lines B and D parallel to line A and tangent to the inflow mass diagram at its maximum and minimum points, E and F. The vertical distance between lines B and D represents the minimum required equalization volume of 30,000 gallons, which is approximately 12.5 percent of the average daily wet-weather flow in this example.

In addition to the volume required to equalize typical wet-weather peaks, the basin must be sized to accommodate any anticipated concentrated plant wastewater streams. Anaerobic digester supernatant and sludge dewatering filtrate are periodically discharged to the front end of the treatment plant and usually have higher organic and nutrient concentrations than typical municipal wastewater. Due to their periodic discharge, these flows create shock load conditions which reduce plant efficiency. COD and ammonia concentrations of 10,000 to 20,000 mg/l and 1,000 mg/l, respectively, are common and consequently create a high oxygen demand. Equalization of these loadings is extremely beneficial to overall plant performance.

The following table is an estimate of the equalization basin's volume requirements for the example shown in Figure 3-1:

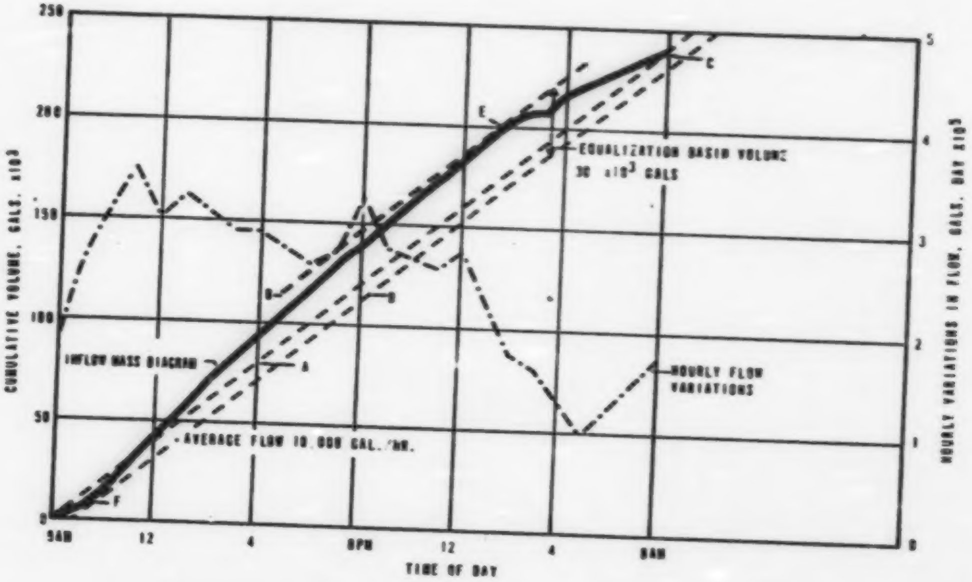
Table 3-1

Equalization Requirements

Source	Range of Equalization Needs percent of flow
Wet Weather Flow Equalization	12.5
Digester Supernatant	0.3 to 1.4
Sludge Dewatering Filtrate	0.5 to 1.5
TOTAL	13.3 to 15.4

The maximum volume requirements to equalize wet-weather flows will depend on the magnitude of the infiltration and extraneous surface water that enters the wastewater collection system. In some cases, it may not be economically feasible to equalize extreme

FIGURE 3-1
EQUALIZATION REQUIREMENTS FOR A TYPICAL WET WEATHER FLOW



peaks of wet-weather flow. Regulatory requirements involving prohibition of treatment plant bypassing, however, will favor the construction of some type of equalization facility. Examination of the plant's past flow records will facilitate the selection of a particular wet-weather flow for design purposes.

For successful operation of equalization basins, mixing and aeration of the wastewater are required. Mixing is necessary to prevent deposition of solids in the basin, and aeration is required to prevent septicity.

A typical equalization flow schematic is shown in Figure 3-2. The flow enters the equalization basin by gravity, and the basin contents are pumped to the primary treatment units using continuous-flow, variable-speed pumps. The maximum pumping capacity of the equalization pump station should be sufficient to handle the maximum flow expected, even though the equalization basin may be designed to equalize a somewhat smaller flow.

3.3 Process Designs and Cost Estimates

Capital cost estimates, prepared for equalization facilities for plants having capacities of 1, 3, and 5 mgd, are shown in Table 3-2. These costs do not include land costs, contingencies, engineering design, and bonding.

Table 3-2

Capital Costs for Equalization Facilities
(ENR Index 1,500)

Plant Size mgd	Capital Costs for Equalization Facilities In Thousand Dollars
1	$210
3	450
5	600

The above costs are based on the typical flow diagram shown in Figure 3-2. The volume of the equalization basin was based on 15 percent of the treatment plant daily capacity.

The basin dimensions should be selected to avoid interference between aerators and to minimize fluctuations in basin water level. Aerator manufacturers recommend a minimum basin size of 15 to 50 feet square and a minimum depth of 5 to 8 feet, depending on the particular aerator horsepower used. To avoid large volumes of dead storage in the equalization basin as a result of aerator operating requirements, the use of a compartmented basin is suggested. For example, two compartments of a four-compartment basin could be used to equalize diurnal flow variations, while all four compartments could be used to equalize wet-weather peaks. When floating surface aerators are used, care must be exercised to maintain a minimum water level to protect the aerator. This may be

CHAPTER 14

OPERATION AND MAINTENANCE REQUIREMENTS FOR UPGRADED PLANTS

14.1 General

The discussions in the preceding chapters of this manual demonstrate the fact that upgrading existing wastewater treatment facilities involves the consideration and possible inclusion of a multiplicity of processes and process equipment. Operation and maintenance requirements for a plant incorporating such a diversity of processes are frequently complex. Since this manual is directed principally to the process design of upgraded facilities, it is not appropriate to the scope or purpose of this document to present a detailed operation and maintenance manual. Rather, some guidelines and pertinent discussion will be presented to assist in the development of an operation and maintenance program for an upgraded plant.

14.2 Responsibility of the Design Engineer

One of the first and more obvious considerations to the engineer faced with upgrading existing wastewater treatment facilities is the need to consider current plant operation and maintenance, either as causes of problems requiring upgrading or as means of accomplishing the required improvement in plant performance. As has been discussed in the preceding chapters dealing with individual unit processes and operations, an important first step in upgrading plant performance is to review operational and maintenance practices, and the adequacy and competency of plant operating staff and supervision. It is frequently possible to effect a significant improvement in plant performance, without the need for capital expenditures, by modification of plant practices, by addition of operating and/or maintenance staff, by improved or supplemented supervision, or by training of existing staff.

In September, 1970, the Federal Water Quality Administration (now EPA-OWP) issued "Federal Guidelines - Design, Operation and Maintenance of Wastewater Treatment Facilities" (1). The latter section of this document includes "Guidelines for Operation and Maintenance". Projects for which Federal grant assistance is requested are expected to comply with these guidelines as well as with technical bulletins which the Environmental Protection Agency will be issuing from time to time. The aforementioned guidelines provide information on the basic minimum requirements for federally assisted projects on such matters as personnel, records, reports, laboratory control, and process control, and present a suggested guide for an operation and maintenance manual.

The design engineer has the responsibility to be fully knowledgeable of the operational factors of the plant for which he is responsible and to relate that knowledge to the alternative means of upgrading the plant's performance. For instance, in both the selection and evaluation of alternative upgrading measures, he must consider total annual costs,

This copy taken from the "Process Design Manual for Upgrading Existing Wastewater Treatment Plants" prepared by Roy F. Weston, Inc. under Contract 14-12-933 with the EPA Technology Transfer Program.

including operation and maintenance, as well as capital construction costs. Some guidelines to assist in such comparisons have been included in the previous discussions on unit processes.

The design engineer should make it a point to obtain all pertinent plant operating records and to solicit the observations, comments, and suggestions of the operating staff so that optimum use may be made of the experience and knowledge of the particular plant and system. He should also prepare or assist in the preparation of operation and maintenance manuals for the upgraded plant, and in the on-site training of operating, maintenance and supervisory personnel to ensure that the function, capabilities, and limitations of the upgraded facilities are adequately communicated to those responsible for their operation. He should be retained to remain involved during at least the initial operational period to utilize most effectively his expertise and knowledge of the process.

The design engineer also has the responsibility to assess the adequacy of available information on existing plant performance, to define the additional data required, and to develop and direct or execute a program to obtain such information. A program to obtain adequate design information can range anywhere from simple qualitative observations through quantitative samplings and analysis programs to laboratory or pilot-scale treatability studies. This charge presents a very real challenge to his professional judgement, requiring the balancing of the constraints of economics, timing, and staff availability against the need for a fully reliable and effective treatment facility. The information presented in this manual provides direction and guidance to assist the design engineer in the exercise of this professional judgement.

14.3 Instrumentation and Automatic Operation

Historically, wastewater treatment plants have not incorporated sophisticated instrumentation or automated operation. However, with the advent of improved instrumentation and the need in many cases to consistently maintain plant effluent quality, the use of such systems may be indicated.

14.3.1 Instrumentation

In general, instrumentation leads to improved performance, increased efficiency, and reduced operating costs. It also offers permanent records for system evaluation, and generates data for required State regulatory reports. There are three basic types of parameters which may be monitored in wastewater treatment: mechanical, physical, and chemical. Table 14-1 contains a summary of instrumentation available for use in wastewater treatment plants (2). Some common on-line instruments which have application for various unit processes in a conventional wastewater treatment processes follow.

Table 14-1

Wastewater Treatment Instrumentation

Primary Measuring Method	Metering Function	Some Applications
Pressure	Pressure Level	Pump discharge pressure Force main pressure Wet-well level Chemical storage level High-water alarm Gas storage pressure
Mechanical position or motion	Level Position	Digester cover level Valve position indication High-water alarm
Electric and electronic	Temperature Flow Level	Incinerator temperature Wet-well overflow Diversion overflow Digester temperature Magnetic flow metering
Photoelectric	Turbidity Total carbon analyzer	Final effluent turbidity BOD
Differential pressure	Flow Ratio control	Wastewater flow Air flow Cl_2 flow Treatment chemical flow Flow proportioning
Radiation absorption	Specific gravity Density Moisture content Level	Sludge density control Wet-well level Sludge-well level Filter cake moisture content
Electrochemical reaction	D.O. pH ORP Cl_2 concentration Conductivity	Aeration tank D.O. Digester pH Final effluent pH Cl_2 residual Effluent D.O.

Source: Salvatorelli, JWPCF (2)

14.3.1.1 Bar Screen Control

Instrumentation for bar screen control should include an inlet level indicator which actuates the cleaning mechanism when the water reaches a predetermined level in the influent chamber. In addition, a high water surface level alarm and a high torque alarm in the bar screen cleaning mechanism should be considered.

14.3.1.2 Grit Chamber Control

Grit chamber controls should include a high torque alarm on motor controls and perhaps a timer on grit removal mechanisms. Both the bar screen and grit chamber high torque alarms may be connected to a high torque motor shut down.

14.3.1.3 Raw Wastewater Flow Measurements

Some means of measuring wastewater flow include flumes, venturi meters, orifice plates, and magnetic flow meters. Flow measurement is required to provide a record of influent flow to evaluate plant and unit process efficiencies and to dispense chemicals.

14.3.1.4 Primary Clarification Control

One method of measuring primary sludge is with a magnetic flow meter. Self heating meters are available and quite useful where grease is a problem. Sludge depth can be indicated using either infra-red adsorption, photo-electric opacity, ultra-sonic amplitude detection, or radiation amplitude detection. Sludge pumping can then be initiated automatically and controlled by a predetermined timing operation or by using a sludge density meter in the clarifier underflow line.

14.3.1.5 Secondary Clarification Control

The motor controls of the secondary clarifier, like the primary clarifier, should be connected to a high torque alarm and motor shut down switch. Control of secondary sludge drawoff must be considered in conjunction with subsequent sludge dewatering equipment. The control of secondary sludge is much more sensitive than control of primary sludge.

Sludge blanket depth can be determined by the same means as used to determine primary sludge depth. Present measuring and sensing equipment is not as sensitive as required for most secondary sludges. A measurement of sludge blanket depth with a predetermined timed sludge drawoff may prove to be the optimum instrumentation. Control of recycled sludge is probably best accomplished by utilizing a magnetic flow meter as the flow element. A conventional flow element can be used, but would be recommended only if supplemented with a continuous purge system, inspection access, and a good maintenance program. The control loop should include the flow element and controller with proportional and reset controls. Waste sludge should likewise be metered and recorded.

14.3.1.6 Dissolved Oxygen Control (Activated Sludge)

Reasonably reliable and accurate dissolved oxygen monitoring equipment has been developed and field demonstrated in recent years. Automatic control of dissolved oxygen (D.O.) in activated sludge plants is generally more useful and justifiable on the basis of economics of operation rather than on reliability or quality of performance. That is, automatic D.O. control can be used to more effectively match power input to oxygen demand than to upgrade biological removal kinetics. Placement of the D.O. probe to obtain a true average reading is a difficult problem Assistance from the probe manufacturer should be requested if necessary.

14.3.1.7 Trickling Filtration Systems

Trickling filtration generally requires less sophisticated instrumentation than does activated sludge. The instrumentation for trickling filters generally includes control of effluent and sludge recycle depending on the individual flow diagram. Control versatility must be provided if a two-stage plant is to be operated in series and parallel.

14.3.1.8 Chlorination Systems

The chlorination system is generally controlled with a final residual analyzer. The system should include recording as well as control equipment.

14.3.1.9 Sludge Handling Systems

Instrumentation packages for sludge handling systems, including air flotation, anaerobic digesters, vacuum filters, and centrifuges, are often supplied by the individual manufacturer. Coordination is required to couple control to input and output systems, and to relay required information to a central control panel, if used. Detailed discussion of the specific instrumentation involved is omitted here because of the many approaches which may be applicable.

14.3.1.10 Level Control

Level control may be critical in the operation of various unit processes. High and low level alarm may be coupled with pump control as required. Pneumatic bubbler sensing systems offer the simplest approach to level control. A water system that creates an air pressure suitable for a level sensor is now available and is quite useful in systems where there is no instrument air available.

14.3.1.11 Foam Detection

Foam levels can be detected by use of photoelectric cells which use a modulated output. The output from the foam detection device can then actuate foam control equipment.

14.3.2 Automated Instrumentation

Presently there are automated wet chemical procedures for COD, ammonia, phosphates, and other specific materials. In addition, combustion procedures for total carbon (TC), total organic carbon (TOC), and total oxygen demand (TOD) have been developed and applied successfully. These automated analytical systems provide levels of sensitivity which are not normally obtained in manual operation of the same procedures, and in some cases can be used for process control.

14.3.3 Centralized Control Panel

The use of a centralized control panel should be considered in any fully instrumented system. Instant data availability at a central control point permits immediate reaction to alarm conditions and, with proper manual standby controls, gives the experienced operator complete control over most situations.

14.3.4 Summary

The decision to utilize sophisticated instrumentation and/or automatic controls depends, among other factors, on plant size and the importance of consistently maintaining effluent quality.

There is a scarcity of information regarding the current investments made in controls for wastewater treatment plants. Salvatorelli (2) reported that approximately 4.5 percent of the total capital cost of a recently constructed municipal wastewater facility was invested to completely monitor and control the facility. A somewhat analogous comparison is the 12 percent invested by chemical and petroleum industries in 1968 for manufacturing process controls (3). Andrews (3) reported that, based strictly on an economic analysis, a 9 percent investment in instrumentation and controls for a 100 mgd activated sludge plant could be justified if the plant performance could be improved from 87 to 92 percent BOD removal. Smaller plants (less than 5 mgd) may not be able to justify expenditures for such instrumentation and controls even if plant performance could be improved as above. The possibility should be analyzed carefully and weighed against other alternatives, e.g., better use of the money through additional treatment capability.

14.4 Operation and Maintenance Requirements

Operation and maintenance requirements (and their associated costs) are generally defined by three factors: chemicals, required power, and labor.

14.4.1 Chemicals

Chemical quantities and their associated annual costs can be readily estimated from required dosage rates (average for annual costs, maximum for storage and feed rate requirements), wastewater flow rates (average and maximum considerations as with dosage), and unit costs appropriate to the particular chemicals used. The designer should bear in mind that the form in which the chemical is received, stored, or fed can have a significant impact on the amount of operating labor required in its handling and feeding, as well as on the initial construction costs of the handling and feeding facilities. This is an individual consideration requiring evaluation of relative costs as delivered for the various forms, the availability of existing storage, handling, and feeding facilities, and staffing policies.

14.4.2 Power

Incremental power requirements and costs are likewise relatively simple to estimate, being the function of additional connected horsepower, its operating frequency and duration, and unit power costs. Local utility policy with respect to demand rates, of course, must be included in the economic comparison. In addition, recent recommendations by EPA state that the treatment facility should be capable of satisfactory operation during emergencies and power failures (1). To achieve this degree of reliability, duplicate sources of power for essential plant elements must be provided.

14.4.3 Labor

EPA will be issuing recommended wastewater treatment plant staffing guides in the near future. These guides will offer the engineer a means of more accurately estimating manpower requirements by establishing a step-by-step methodology for itemizing those tasks which must be performed at treatment plants, and for applying related manpower needs to these tasks. One step further is the need for employing personnel who are qualified to adequately and efficiently operate the wastewater treatment plant. Most states currently have mandatory operator certification programs, while the majority of the others have voluntary operator certification programs.

14.4.3.1 Operating Manpower Requirements

The first step in the development of estimates of additional operating manpower requirements for upgraded facilities is a thorough analysis of existing operations to establish the adequacy of present staff and operational procedures. This must be done whether the upgrading procedure selected involves the modification of operating procedures or requires the addition or modification of unit processes. The analysis should account for specific local conditions, problems, and objectives.

Following this analysis of existing wastewater treatment facilities, the engineer must make a judgement as to the efficiency of operations and the qualifications of the staff in relation to performance of upgraded facilities. Factors effecting such a judgement include:

1. The physical arrangement and geographic dispersion of the system.
2. The relative complexity of the present versus the upgraded facilities.

3. **The level of training and experience of the present staff in relation to the level required by upgraded processes.**

4. **The relative degree of laboratory and analytical control required for the upgraded versus the existing facilities.**

The operational analysis and estimate of manpower requirements should be done on an individual basis for each plant being considered for upgrading. Manpower requirements should be estimated and recommended as dictated by the specific local conditions.

14.4.3.2 Maintenance Manpower Requirements

Similar to the development of estimates for additional operating manpower requirements for upgrading facilities, the first step in the estimation of additional maintenance manpower requirements is a thorough analysis of existing maintenance staff and procedures. The development and implementation of an appropriate preventive maintenance program for wastewater treatment facilities alone can represent a significant upgrading procedure to the extent that it provides for reliable, consistent plant operation and performance. Again, the design engineer responsible for facility upgrading must exercise his judgement as to the sufficiency of maintenance programs and the adequacy of the qualifications and performance of the maintenance staff. This information must in turn be related to the requirements of the upgraded processes and facilities. The factors influencing the engineer's judgement on these aspects of facility performance are similar to those discussed previously on the topic of operation requirements.

In the estimation of total maintenance costs for upgraded facilities, the engineer must include maintenance materials and supplies in addition to maintenance labor. Again, he is usually dealing with incremental costs and must assess the adequacy of present lubrication practices, spare parts stocking and replacement practices, and similar maintenance procedures involving expendable materials and supplies.

14.5 References

1. *Federal Guidelines - Design Operation and Maintenance of Waste Water Treatment Facilities*. Federal Water Quality Administration, September, 1970.

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3. Andrews, John F., *Control of Wastewater Treatment Plants - the Engineer as an Operator*. Water and Sewage Works, 118, No. 1, pp. 26-32 (1971).

SLUDGE DISPOSAL BY LANDSPREADING

I INTRODUCTION

A Wastes removed from water must be disposed of in air, water, or land. Recycling of some components may delay disposal.

B Fresh water is no longer acceptable as a receptacle for wastes.

C Ocean disposal of organic waste and some chemicals is a questionable practice.

D For many cities and towns spreading liquid wastewater sludge on the surface of the land is the simplest, most inexpensive method of sludge disposal. It conserves plant nutrients and organic matter for utilization by plants.

E There are potential hazards involved with improper landspreading practices. Recycling nutrients and water in soils is beneficial if correctly applied. Certain hazardous or damaging materials accompanying the nutrients are serious problems.

II HISTORY

A Disposal of wastes by returning them to the soil is an ancient practice that is still popular.

1 Bunslau, Prussia irrigated with sewage in 1559.

B "Nightsoiling," the practice of returning untreated human wastes to the soil, has been important to the economy of many countries by maintaining soil productivity, but it has led to an association of unsanitary conditions with land disposal.

C Land disposal of wastes has enjoyed flourishes of popularity followed by periods of disfavor. We are now in a period of increasing interest.

III THE SOIL AS RECEPTABLE FOR WASTES

A The soil is a natural body at the earth's surface that supports or is capable of supporting plant growth. It is made up of mineral material, organic matter, biological populations, solutions, and air. It is dynamic. Volumetric composition of a typical topsoil:

Mineral matter - 45%
Air and solution - 50%
Organic matter - 5%

B A typical soil has properties which equip it to serve as an assimilator of wastes. It performs the following functions:

Filters
Buffers
Adsorbs
Transforms chemically
Transforms biologically
Supports plants that reuse nutrients contained in wastes

C Soil properties usually desirable for waste assimilation

1 Deep
2 Permeable
3 Fine textured enough to sorb ions
4 Level (except for spray runoff systems)

IV THE SLUDGE

A Digested sludge is a good fertilizer

1 Characteristics of sludge are determined by composition of wastewater, treatment processes, and the sludge conditioning and handling techniques used.

2 A suspension of organic and inorganic solids removed from wastewater.

3 Liquid sludge is high in water content. The water content varies from 90% in chemically precipitated organic sludge to 99% in activated sludge.

4 Before stabilization it contains living pathogens and putrescible organic solids.

5 Stabilization and sometimes disinfection are needed before spreading it on land.

6 It contains all essential elements for plant nutrition. Potash is low in comparison to nitrogen, and phosphate.

7 Dried sludge is not competitive with inorganic fertilizers, but a two-inch application of liquid sludge from Chicago will supply the nutrients for an average corn crop in Illinois (Dalton, 1967).

8 Fertilizer value of dried sewage sludge: (Anderson, 1955)

	Activated	Digested
Nitrogen (Average 1951-55)	5.6% of solids	2.4%
Phosphate	5.7%	

Average fertilizer value of liquid digested activated sludge from Chicago:

	% of solids by weight
Nitrogen	5
Phosphate	3
Potash	1/2

B Sludge is a good soil conditioner

1 Organic sludge matter is similar to soil humus when it is digested.

2 Sludge improves soil structure, permeability, water-holding capacity, and capacity to hold nutrients.

3 Water and other constituents of sludge may promote plant growth.

C Sludge must be stabilized before it is spread on the land.

1 Anaerobic digestion is the most popular method.

a It reduces volatile solids leaving a stable humus-like organic portion.

b Pathogens are reduced.

c Control of digesters is difficult. Poorly digested sludge is not suitable for land spreading.

2 Aerobic stabilization is a very promising method of stabilizing sludge for land spreading.

a It produces a brown, almost odorless, nitrified sludge that is very stable.

b It removes grease and reduces pathogens.

c Aerobic stabilization is popular with many small treatment plants.

d Power is required for aeration.

e Aeration is difficult with primary sludge, but easier with biological sludge.

3 Chemical stabilization

a Liming to raise pH to 11.0 for 24 hours has been effective in stabilizing sludge and killing pathogens in the laboratory. Lime sludge has promoted growth of grass in short-term greenhouse studies. Long-term results are expected to be beneficial to soils.

b Sulfur dioxide stabilized sludge in the laboratory. The sludge has not been tried on the soil but is expected to have a slight acidifying effect that can be controlled with limestone.

c Other chemicals applied to wastewater or sludge--aluminum and iron salts or organic polymers have had no detrimental effects in short-term studies. The soil clogging effect produced by the gelatinous aluminum and iron sludges is expected to be offset by the improved physical condition resulting from increased organic matter in the soil.

V WAYS OF DISPOSING OF SLUDGE BY LAND SPREADING

A Objectives include:

1 Disposal of as much sludge as possible without degrading the environment or creating a nuisance;

2 Improving low quality land;

3 Fertilize vegetation and improve soils;

4 Combinations of the first three.

The objective chosen will depend upon the kind and amount of land available, the kind and amount of sludge available, public attitude toward sludge spreading, and personal preference of officials.

B Conveyance

1 A very popular way for small treatment plants to dispose of sludge is to haul directly from the digesters to cropland or forest, and spray from a tank truck. Large volume or long hauls are expensive.

2 Barges have been used by coastal cities to transport sludge for ocean disposal. They have also been tried at Chicago. A large volume of sludge and navigable water are requirements for feasible barging. Barge transportation offers some flexibility, but sludge thickening would be needed. It also requires loading and docking facilities at the plant and docking, unloading, and distribution system at the spreading site.

3 Railroad transportation can be used if rail lines are conveniently located for the spreading site.

4 Pipelines are economical and dependable for large cities to transport sludge up to 150 miles or more. Capital costs are high but operating costs are low when compared to the other three methods.

5 Sludge is most frequently conveyed to land disposal sites by tank truck from small treatment plants and pipelines by large treatment plants.

C Application techniques

1 Irrigation methods are usually used when the objective is fertilization of vegetation.

a Ridge and furrow irrigation requires level land or grading to achieve contour furrows. It is better adapted to cold weather operation and it offers less aesthetic problems than spray irrigation.

b Spray irrigation is flexible and requires less soil preparation. It can be practiced on sloping land and with a variety of vegetative cover. The rainbird is a common type. Aerosol drift may occur.

c Flood irrigation usually requires land shaping.

d Injecting sludge into soils or plowing it in with a moldboard plow gets it out of sight and prevents odors. The methods are not adapted to continuous applications on cropland.

e Flood and till is used when crops are not on the land.

2 Disposal or reclamation of low quality land is usually done by some variety of flood and till methods.

D Loading rates -- depend upon the system -- fertilization, reclamation or disposal.

1 The proper annual application rate also depends upon sludge composition, concentration of solids, soil characteristics, climate, vegetation, and cropping practices.

a Reclamation applications up to 1000 tons per acre per year have been used successfully.

b Soil conditioner and fertilizer applications have ranged from 0.5 to 300 tons of solids per acre. Burd suggested 100 tons per acre for cropland. Lunt (1961) suggested 10 tons of solids per acre of sludge containing 2.5% nitrogen as an optimum for crops grown on medium or sandy textured soils. A general rule-of-thumb is:

Up to 100 tons per acre per year for ideal conditions
10-25 tons per acre per year for average conditions
< 5 tons per acre per year for adverse conditions

VI FACTORS DETERMINING SAFE SLUDGE DISPOSAL ON LAND

A Nature of sludge

1 Sludge containing high concentration of heavy metals or other toxic substances should not be applied to land. Some heavy metals and minor elements are toxic to plants at relatively low concentrations. Only on acid soils and with heavy applications has toxicity been experienced. Management to precipitate metals in soils may be needed.

2 High nitrogen content may lead to nitrate pollution of aquifers or surface water. Crops may remove up to 300 lbs. of nitrogen/A/-per year. Nitrogen removal may be necessary where heavy applications of sludge are made.

3 Hygienic conditions

a Digested sludge contains some living pathogens.

b Most pathogens die quickly in soil, but some may live for several weeks.

c Fruits or vegetables which grow in contact with the soil and are eaten raw should not be fertilized with sludge unless it has been disinfected.

d Disinfection may be done economically with heat, chemicals, or storage in a lagoon. Pasteurization is practiced in Europe.

e Most soils are effective filters. Water that has percolated through five feet or more of soil has been found to be free of pathogens.

B Shallow, steep, or coarse textured soils and runoff should be avoided. Wet, poorly drained soils are undesirable.

1 Erosion should be controlled to prevent sludge from washing into surface water.

C Loading rate

1 Overloading the soil can cause odors, insects, and aesthetic problems. Well stabilized sludge applied in thin applications has caused neither odor nor insect problems.

D Land use determines the precautions required.

1 Sludge applied to land that is isolated and inaccessible to the public may not require the same degree of disinfection and stabilization as sludge that is applied to recreation land or to other land accessible to the public.

2 Some cities require disinfection of sludge that is to be spread on pastures. However, spreading digested sludge on pastures has been widely practiced in this country without known outbreaks of disease.

VII ECONOMICS

A The cost of spreading sludge on land is less than for other forms of disposal for most cities and towns.

1 The net cost of drying and selling sludge as fertilizer at San Diego was $35.00 per dry ton of solids in 1951 to 1956 period. In 1959 and 1960, the cost of spreading liquid sludge on the land was $9.91 per ton (Burd, 1968).

2 Liquid sludge spreading to make productive soil out of sand at New York City was reported to cost about the same as barging to the ocean, $7.50 per ton.

3 Chicago's plan developed in 1967 compared costs of landspreading with the methods they were using and with incineration.

		Gross
Drying for fertilizer	$45.00 net	$60.00
Wet air oxidation	50.00	
Digestion and permanent lagoons	49.00	
Digestion and reclamation of farm land	15.00	
Digestion and reclamation of strip mines	16.00	

4 It is easily the most economical method for some small treatment plants. Others find drying beds with air dried sludge picked up by farmers and gardeners more economical.

VIII OBSTACLES TO SLUDGE SPREADING ON LAND

A Public resistance to accepting waste from the city on rural land forced revision of plans at Chicago, Charleston, West Virginia, and Columbus, Ohio.

1 In contrast, a number of cities charge a small fee for liquid sludge.

2 Many farmers in Pennsylvania desire more sludge than is available to them.

B Sludge spreading has no glamor or profit for the designer.

C Design criteria is lacking. Guidelines are available in the form of reports of research and successful land spreading projects, but precise design criteria is not.

D Land, whether purchased or obtained under agreement, may be difficult to get without an effective educational program.

E There is a general misconception that near many of the larger cities no open land exists that would be suitable for sludge spreading. Satisfactory land exists within about 100 miles of every city in this country.

F There are unfounded concerns about disease hazards, odors, insects, and toxic substances when sludge is applied to the soil.

G Past failures caused by unstable sludge, poor land management, and overloading of soil have tended to prejudice people against land spreading.

1 Farming should be done by farmers.

IX SOME CITIES THAT USE LAND SPREADING

A Manassas, Virginia

1 Has a 2 MGD activated sludge plant.

2 Digested sludge is trucked about three miles and spread on farm land from back of truck.

3 Pasture, hay, and some corn and small grain are grown.

4 Rate of application is about 10,000 gallons per acre per year--less than enough to meet the nutrient requirements of the crops.

5 Land spreading was adapted when alum sludge failed to dry on drying beds.

B San Diego, California

1 Mission Bay Park was created by spreading digested sludge on dredged sand from shallow bay.

2 A seven and one-half mile long 8" pipeline is used to transport sludge to Mission Bay Park.

3 Up to 1000 tons per acre of sludge has improved soil for grass and ornamental shrubs.

4 New low odor thresholds caused by luxury hotels and park use caused adoption of present drying bed technique instead of flow and till system formerly used.

5 Sludge from pipeline is spread to a depth of about 8" on 15 acres of drying beds of one and one-half acres each.

6 Supernatant from drying beds is decanted in series. Cultivation starts as soon as sludge is dry enough.

7 Dried sludge is removed from beds and stockpiled for use in parks.

8 Cost of building parkland with sludge is about 1/4 that of importing topsoil.

C Oceanside, California

1 Digested sludge is hauled by truck and dumped into a trench 3' deep and 18" wide. Although the sludge has been buried in the same area for eight years, they never see sludge in the sand. Microbial oxidation is rapid.

D Chicago, Illinois

1 The city produces about 1000 tons per day of sludge solids.

2 Sludge has been handled in four ways.

a Wet air oxidation

b Heat drying for fertilizer

c High rate digestion--lagoon storage

d Imhoff

3 Plans call for high rate digestion followed by disposal on low grade farmland.

4 Pipeline transportation was determined to be the most economical with barging second.

5 After heavy applications to build up productivity, cropland will receive 20 tons of solids per year.

6 Nitrogen has proven to be the element in sludge that will first limit sludge applications. Heavy metal accumulations have not been excessive after four years of applications.

7 Adverse public reaction has prevented full implementation of the land spreading program.

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